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Project Manual for Full-Scale Dual Alkali Demonstration at Louisville Gas and Electric Company Preliminary Design and Cost Estimate

Louisville Gas and Electric Company, Ky.

Prepared for

Industrial Environmental Research Lab., Research Triangle Park, N.C.

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Industrial Environmental Research Laboratory
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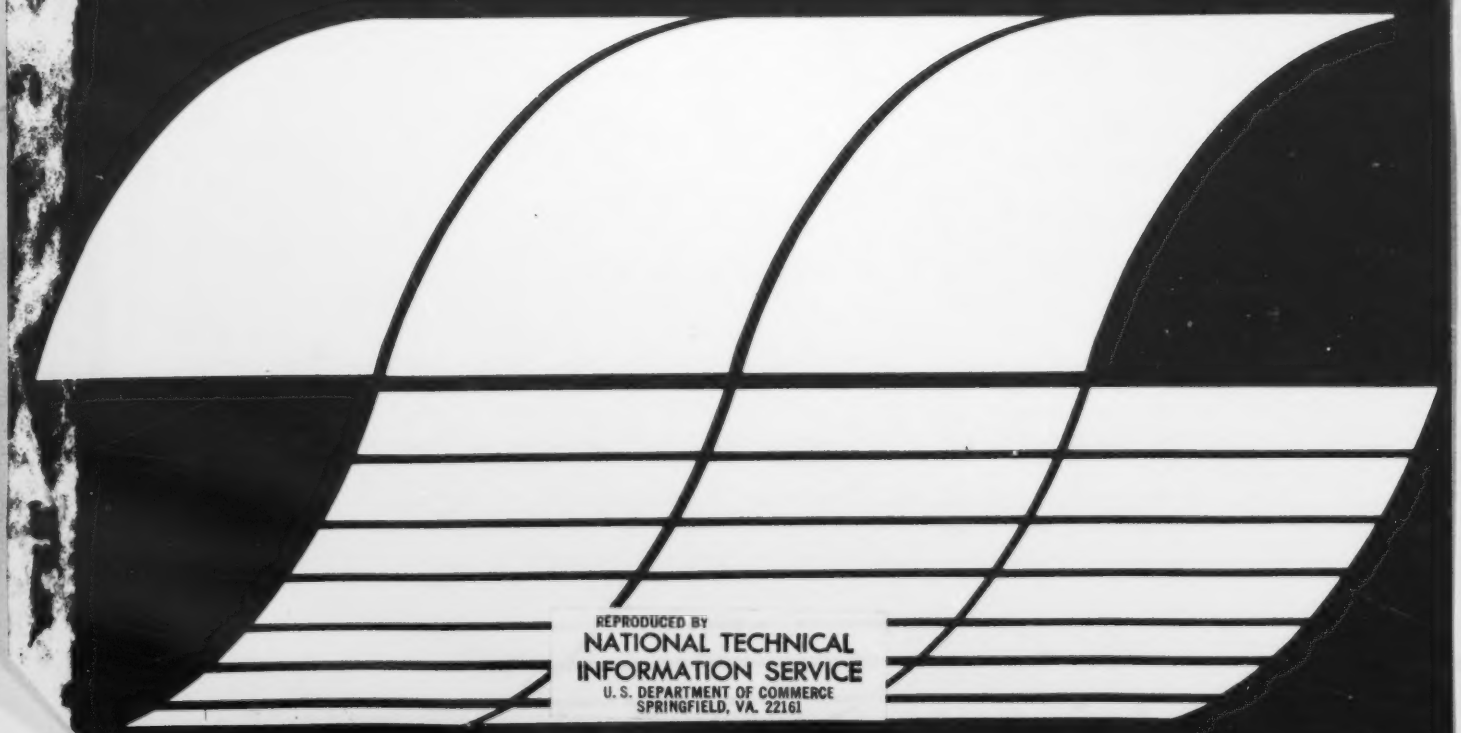

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16. ABSTRACT The report is the project manual for the dual-alkali system, designed by Combustion Equipment Associates, Inc./Arthur D. Little, Inc. and being installed to control SO2 emissions from Louisville Gas and Electric Company's Cane Run Unit No. 6 boiler. The project consists of four phases: I--preliminary design and cost estimates; II--engineering design, construction, and mechanical testing; III--startup and performance testing; and IV--1 year operation and testing. Developed as part of Phase I, the project manual includes a detailed description of the dual-alkali process chemistry, the design of the demonstration system at LG and E, material and energy balances for the system, specifications of major equipment items and offsites, and capital and operating costs. Costs for this application have been generalized for new applications on 500 and 1000 MW high-sulfur coal-fired boilers.

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PROJECT MANUAL FOR FULL-SCALE DUAL ALKALI DEMONSTRATION AT LOUISVILLE GAS AND ELECTRIC CO. - Preliminary Design and Cost Estimate

by

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ABSTRACT

The dual alkali process developed by Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) is being installed to control SO_2 emissions from Louisville Gas and Electric Company's (LG&E) Cane Run Unit No. 6. The Federal Environmental Protection Agency (EPA) has selected this system as the demonstration plant for dual alkali technology and is participating in funding of the operation, testing, and reporting of the project. The project covers the full-scale application of the system to Unit No. 6, a high sulfur, coal-fired boiler having a gross peak capacity of 300 megawatts (Mw). The system is expected to start up in late 1978.

The project consists of four phases: Phase I - preliminary design and cost estimates; Phase II - engineering design, construction, and mechanical testing; Phase III - startup and performance testing; and Phase IV - one-year operation and testing.

This report is the Project Manual for the system developed as a part of Phase I. The report includes detailed descriptions of the process chemistry, design of the demonstration system at LG&E, material and energy balances for the system preliminary specification of major equipment items and offsites, and capital and operating costs. The costs for this application have been generalized for new applications on 500 Mw and 1,000 Mw high sulfur coal-fired boilers.

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The principal participating organizations are Louisville Gas and Electric, Inc., Combustion Equipment Associates, Inc., and Arthur D. Little. The persons within each of these companies who were directly involved in the preparation of this report and the work performed during Phase I are listed below:

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R. C. Weeks

Combustion Equipment Associates

T. M. Frank
J. M. Lysaght

Arthur D. Little

I. L. Jashnani (now with Martin Marietta Corporation)
C. R. LaMantia
R. R. Lunt

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APPLICABLE CONVERSION FACTORS

ENGLISH TO METRIC UNITS

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft^2	0.0929 $meters^2$
1 ft^3	0.0283 $meters^3$
1 grain	0.0648 gram
1 in.	2.54 centimeters
1 in^2	6.452 $centimeters^2$
1 in^3	16.39 $centimeters^3$
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal	3.7853 liters
1 Btu	252 calories

I. SUMMARY

A. INTRODUCTION

The dual alkali process developed by Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) is being installed to control SO_2 emissions from Louisville Gas and Electric Company's (LG&E) Cane Run Unit No. 6. The Federal Environmental Protection Agency (EPA) has selected this system as the demonstration plant for dual alkali technology and is participating in funding of the operation, testing, and reporting of the project. The project consists of four phases: Phase I - preliminary design and cost estimates; Phase II - engineering design, construction and mechanical testing; Phase III - startup and performance testing; Phase IV - one year operation and test program. The system is scheduled to start up in October 1978 with the one year test program beginning in early 1979. This Project Manual covers the work in Phase I. A more detailed manual will be prepared at the end of Phase II, and it will be updated at the conclusion of Phase IV.

B. CEA/ADL DUAL ALKALI SYSTEM

The dual alkali system involves absorption of SO_2 using an aqueous solution of alkaline sodium salts. Regeneration of the spent absorbent solution if accomplished using lime which produces a solid waste of calcium-sulfur salts. The process operates in a concentrated active sodium mode* and is capable of SO_2 removal efficiencies in excess of 95% over any range of inlet SO_2 concentration encountered in coal-fired utility boiler applications. In addition to SO_2 removal, the system is highly effective in absorption of chlorides from the flue gas. Particulate removal can also be accommodated by appropriate selection of scrubbers.

The system is operated in a closed loop. The only waste material from the system is a washed filter cake--there are no other liquid or solid purge streams.

C. APPLICATION OF THE DUAL ALKALI SYSTEM TO CANE RUN UNIT NO. 6

The Cane Run Unit No. 6 is a high sulfur, coal-fired boiler having a gross peak capacity of 300 megawatts. The sulfur content of the coal burned in this unit ranges from 3.5% to 6.3% (dry basis); chloride levels typically vary from 0.03% to 0.06% (dry basis). The system design is based on 5.0% sulfur (dry basis) in the coal. Design flue gas conditions downstream of the induced draft fan and at the inlet of the SO_2 control system are as follows:

---

[a] See Glossary for definition of dual alkali terminology.

- Gas flow rate — 1,065,000 acfm
- Temperature — 300°F
- Pressure — -1 to +2 inches W.G.
- SO_2 — 3471 ppm (dry basis)
- Particulate — ≤.0.0537 gr/acf

The CEA/ADL dual alkali system has been designed to meet the following guarantees:

- The emissions from the system shall be no greater than 200 ppm SO_2 (dry basis) at coal sulfur contents of up to 5.0%; 95% removal for coal containing over 5% sulfur.
- The system will cause no increase in particulate matter in the flue gas.
- The consumption of lime will not exceed 1.05 moles CaO per mole of SO_2 removed from the flue gas.
- Sodium makeup will not exceed 0.045 moles of Na_2CO_3 per mole of SO_2 removed from the flue gas.
- The system will consume less than 1.1% of the power generated by the boiler at peak load.
- The filter cake will contain a minimum of 55% insoluble solids.
- The system will have an availability[a] of at least 90% for a one year period.

The principal features of the system at LG&E are:

- The system can be operated from 20% to 100% of the gross peak capacity.
- The flue gas can be reheated by 50 F° at maximum gas flow rate.
- The system is modular in nature and includes two absorber modules, two reactor trains (each train consisting of two reactors in series), one thickener and three filters. The system can be operated with one reactor train at 100% load for short durations, and with one absorber module at up to 60% load.

---

[a]Availability is defined as the ratio of the hours the system is available for operation and the total hours in the operating period (expressed as percent).

- Spare capacity in the system:

Pumps	100%
Filters	50%
Instruments for operation	100%

- The filter cake will be disposed of in an onsite pond after treatment.

- Normally, the system will utilize locally available carbide lime (a byproduct of acetylene production); commercial lime will be used for some test periods.

D. CAPITAL AND OPERATING COSTS

The estimated capital investment and operating costs for the dual alkali system on Cane Run Unit No. 6 and projected costs for generalized systems (500 Mw and 1,000 Mw) are presented in Table I-1 (in 1976 dollars). The generalized systems are based on commercial lime as raw material, 3.5% sulfur in coal, and an annual operating time of 7,000 hours/year. As seen in Table I-1, the capital investment varies from $43/kw to $53/kw, and the operating cost varies from 30¢/10^6 Btu to 36¢/10^6 Btu heat input (equivalent to 2.7 - 3.2 mills/kwh).

TABLE I-1

SUMMARY OF ESTIMATED CAPITAL AND OPERATING COSTS (1976) DOLLARS

Boiler, Mw	300 (Cane Run Unit 6)	500 (—— Generalized	500 ——)	1,000
New or Existing	Existing	New	New	New
Operating Hours	5,256	7,000	7,000	7,000
S in Coal, %	5	3.5	3.5	3.5
SO_2 Removal Efficiency	94.2	78.1 (NSPS)	90	78.1 (NSPS)
Sludge Disposal	Onsite after Treatment	Onsite	Onsite	Onsite
Reheat, °F	0[a]-50	35	35	14
Capital Investment, $/kw	53.2[b]	50.2	52.8	43.7
Annual Cost, ¢/10^6 Btu	23.0[a]-26.7	32.9	35.8	30.3
Annual Cost, Mills/kwh	2.29[a]-2.65	2.96	3.22	2.73

[a] Reheating of the wet gas (from scrubbers) is not in operation (savings in the reheat energy only).

[b] The capital investment for Cane Run Unit No. 6 is equivalent to $53.2/kw based upon gross peak load (300 Mw) and $57.5/kw based upon gross net load (277 Mw).

II. INTRODUCTION

A. PURPOSE OF PROJECT

The project covers the full scale application of the CEA/ADL dual alkali flue gas desulfurization (FGD) system to Unit No. 6, a coal-fired boiler at Louisville Gas and Electric Company's (LG&E) Cane Run Station in Louisville, Kentucky.

The system will be installed on this existing 300 Mw (gross peak capacity) unit to comply with requirements[a] of the Jefferson County Air Pollution Control District, the Kentucky State Division of Air Pollution, and Region IV of the U.S. EPA. EPA has selected the dual alkali SO_2 control process at LG&E as a demonstration system for dual alkali technology and will participate on a cost-shared basis with LG&E for design, operation, testing, and reporting of the demonstration project.

The dual alkali system will be installed following the existing electrostatic precipitator and will have the capability to control the SO_2 emissions to less than 200 ppm dry basis without additional air dilution when burning coal containing up to 5% sulfur. When burning coal containing greater than 5% sulfur, the system will remove at least 95% of the sulfur dioxide in the inlet flue gas. The dual alkali system is not designed for particulate removal; however, it is designed not to increase the particulate loading in the flue gas. As a demonstration system, the purpose of the installation and operation is to establish:

- overall performance - SO_2 removal, lime utilization, sodium makeup, regeneration of spent liquor, water balance, scaling and solids buildup problems, materials of construction, waste cake properties, reliability and availability.

- economics - capital investment and operating cost.

B. SCOPE OF WORK

LG&E will design, construct, and operate (for an operating period of one year following startup) a CEA/ADL dual alkali flue gas desulfurization system on Unit No. 6 at the Cane Run Station, an existing coal-fired boiler. The system shall treat the total flue gas emitted by the nominal 280 Mw boiler. The system shall be capable of treating flue gas equivalent to a minimum of 60 Mw and a maximum of 300 Mw of generating capacity. LG&E shall be responsible for the execution of such subcontracts as are required to accomplish the design, procurement, construction, startup, and operation of the demonstration system.

---

[a] Removal of 85% of the SO_2 present in the flue gas at the scrubber inlet.

The CEA scope includes engineering of the dual alkali system and will provide all the equipment including instruments, control panel, filter building, etc. for the process to LG&E. CEA is responsible for startup and acceptance of the system, and for process guarantees.

The ADL scope includes process engineering help to CEA both during the design and startup of the system, assisting LG&E to conduct a one-year test program, and ADL is responsible for writing reports for the EPA/LG&E contract.

The system shall be designed such that emissions from the stack shall be no greater than 200 ppm SO_2 dry basis without additional air dilution when burning coal containing up to 5% sulfur. When burning coal containing greater than 5% sulfur, the system shall be designed to remove at least 95% of the sulfur dioxide in the inlet flue gas.

The work is divided into four phases:

- Phase I - preliminary design and cost estimate;
- Phase II - engineering, design, construction, and mechanical testing;
- Phase III - startup and performance testing; and
- Phase IV - one year of operation and testing.

This report covers work performed in Phase I of the project. In Phase I, LG&E/CEA/ADL was to prepare a process design and a capital and operating cost estimate for the demonstration system. The process design is for a dual alkali system including system hardware, offsite requirements; and ducting, piping, and other interfaces with the host boiler to render the FGD process fully operational. The term "preliminary design" as used here is intended to cover all work required to adapt the dual alkali flue gas desulfurization process to the particular application and produce a preliminary visualization of all features of the proposed plant in sufficient detail to make a reasonable engineering estimate of the capital and operating costs.

Information generated in preparing the preliminary design was to be assembled in the form of the Project Manual. The Project Manual was to contain sufficient detail to convey the total concept of the proposed plant and to provide a complete basis for the cost estimates. It includes: general process description; material and energy balances; utility requirements; plot plans; offsites; major items of equipment; general arrangement and specification of major equipment; instrument list; piping and instrument diagrams; laboratory and shop capabilities; and process schedule.

A detailed estimate of capital and operating costs for the demonstration plant was to be prepared based on the process design. These costs are broken down into the components used in preparing the estimate.

C. PROJECT SCHEDULE

The overall project schedule covering all phases of the dual alkali demonstration project is given in Figure II-1. The overall project, including the one-year test program, is scheduled for 40 months (with an additional one month for completion of the final draft report).

Phases I and II were scheduled to begin simultaneously to expedite the overall project. Phase I (preliminary design) was scheduled for five months including preparation of the draft report. Phase II (engineering, design, and construction) was scheduled for 24 months starting with the signing of the contract.

A three-month period is allowed for Phase III startup and acceptance testing. The one-year test program (Phase IV) will be planned during the latter part of Phase III. Phase IV is scheduled to start in late 1978 and end in early 1980.

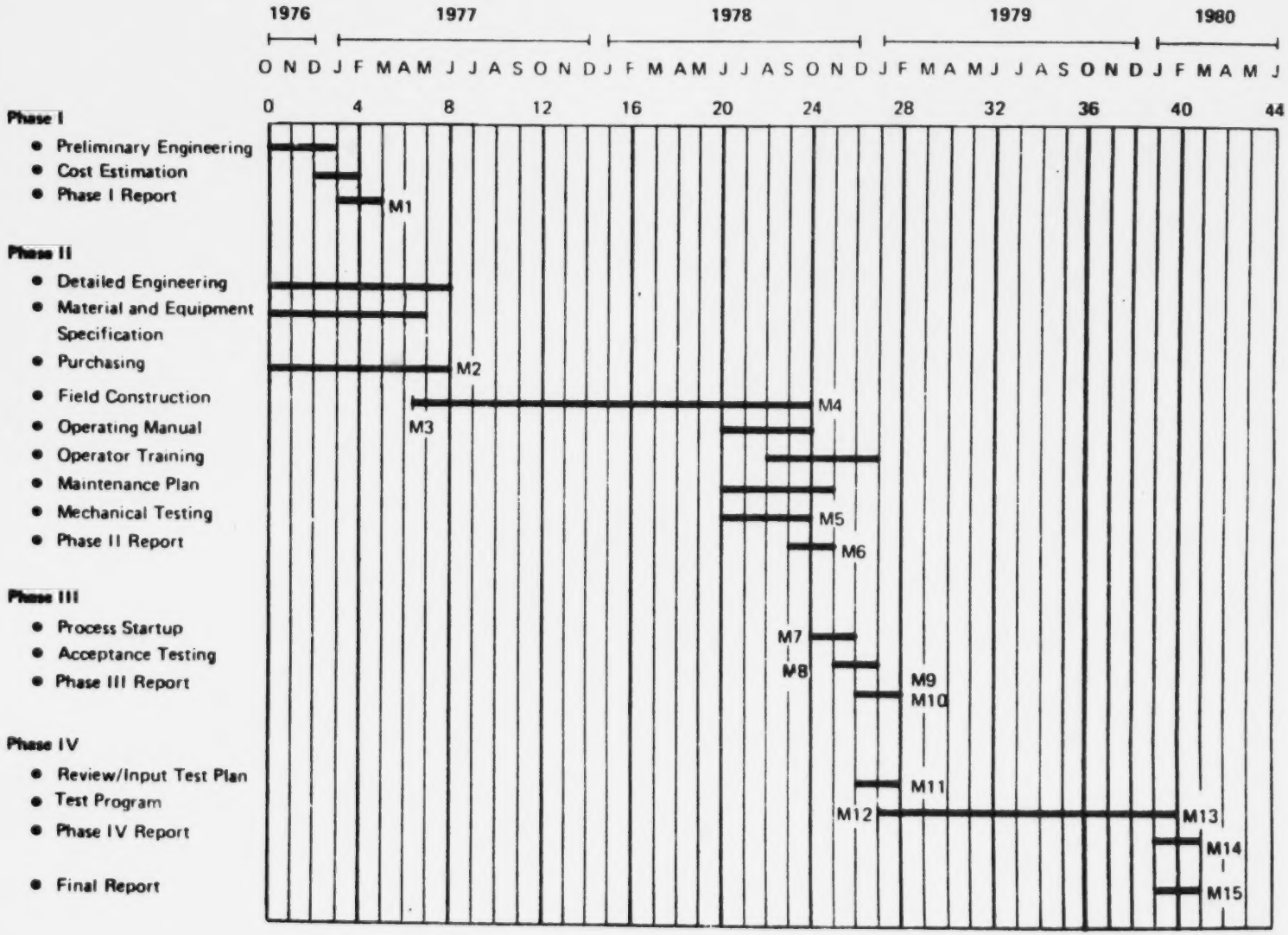

Figure II-1 **DUAL ALKALI DEMONSTRATION OVERALL PROJECT SCHEDULE**

III. CEA/ADL DUAL ALKALI PROCESS

This chapter provides technical information on the process chemistry, process and generalized system design, pollution control capabilities, and waste properties. The SO_2 control system at LG&E differs slightly from the generalized system in this chapter, and is described in the next chapter.

A. PROCESS CHEMISTRY AND SYSTEM DESCRIPTION

The CEA/ADL dual alkali process for SO_2 control from boiler flue gas involves sodium solution scrubbing with regeneration of the absorbent solution using lime.

The system can be conveniently broken down into three process subsystems: gas scrubbing, absorbent regeneration, and solids dewatering.

The equipment utilization and operation of each subsystem depends on specific requirements of the particular application. The following is a general description of the system components for a CEA/ADL dual alkali system operating in the concentrated active sodium mode using lime for regeneration. Figure III-1 shows a generalized flow schematic of the process.

1. Flue Gas Scrubbing

The SO_2 scrubbing system consists of an absorber equipped with a liquid entrainment separator, an enclosed recycle tank to contain the scrubbing liquor, and recycle pumps.

In a generalized system, gas from the electrostatic precipitator (or induced draft fan for the boiler) is forced by a booster fan through the absorber. Gas passes upward through the sprays, a set of trays, and then through a demister. The clean flue gas leaving the tower is finally reheated before being discharged through the stack.

Regenerated absorbent solution, containing sodium hydroxide, sodium sulfite, sodium sulfate and some sodium carbonate, is mixed with scrubber recycle liquor and fed to the top tray of the absorber. The solution flows countercurrent to the gas through the tray system and is collected at the bottom of the absorber in the internal recycle tank. This collected liquor supplies solution both for spraying and the pH control across the trays.

A continuous bleed stream is drawn from the absorber recirculation loop and is sent to the absorbent regeneration system. This bleed rate is controlled on the liquor level in the tank.

The feed forward flow of process liquor to the scrubber system is set according to the pH of the absorber bleed liquor. The flow is adjusted to maintain this pH within a prescribed range dictated by the required SO_2 concentration

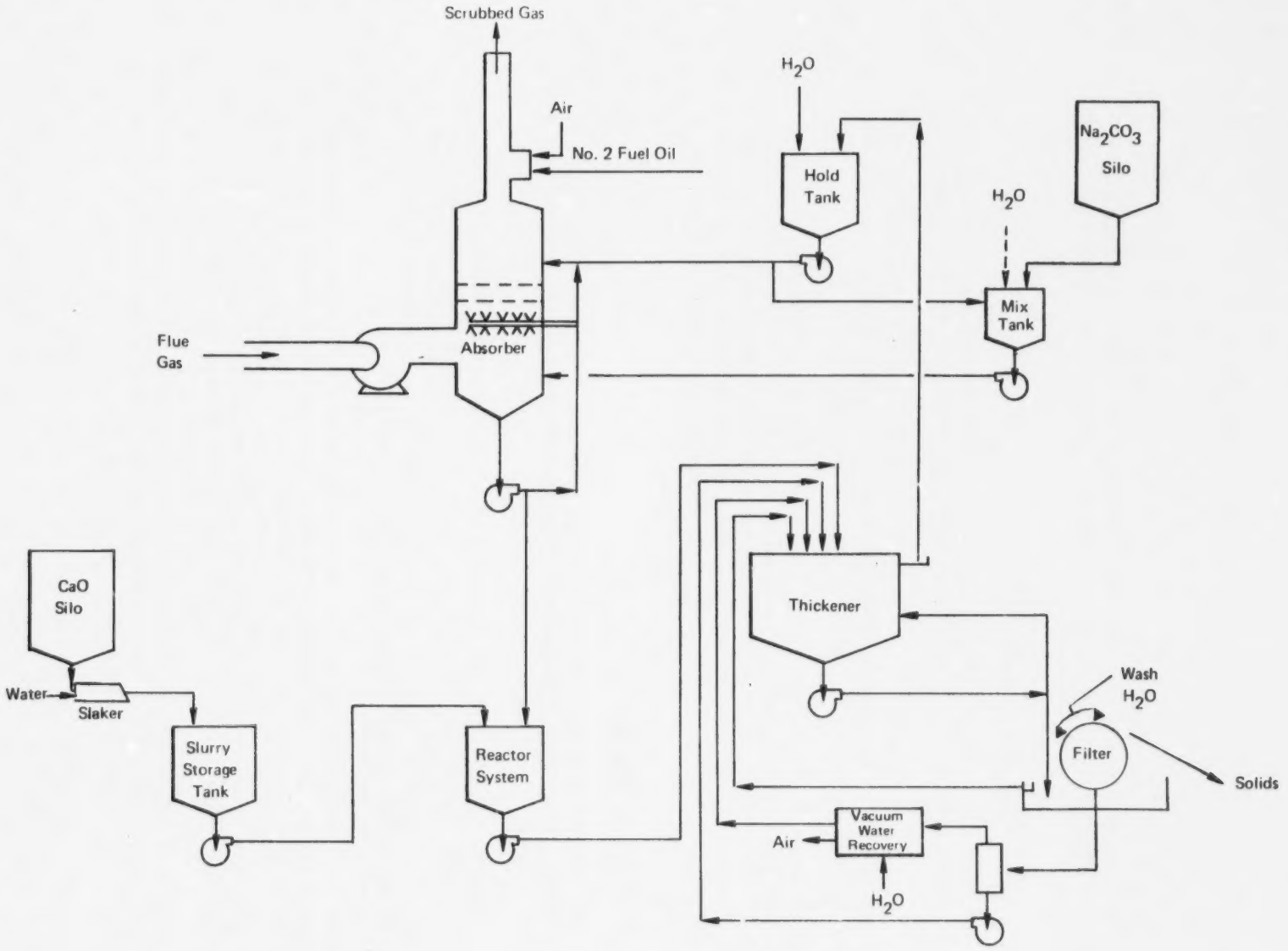

FIGURE III–1 DUAL ALKALI PROCESS FLOW DIAGRAM

in the exit flue gas. For example, a pH of 6.0 may result in the SO_2 concentration (in the exit flue gas) of 200 ppm or lower depending on the operating conditions.

The absorption of SO_2 produces a spent sodium sulfite/bisulfite liquor, as shown in the following reactions:

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O$$

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2\uparrow$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3$$

During absorption, and to a less extent through the remainder of the system, some sulfite is oxidized to sulfate:

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4,$$

converting an "active" form of sodium into an "inactive" form. (The active sodium includes NaOH, Na_2CO_3, Na_2SO_3, and $NaHSO_3$.)

The majority of oxidation occurs in the scrubber system. The level of oxidation experienced is generally a function of the scrubber configuration, oxygen content of the flue gas, and the scrubber operating temperature. At excess air levels (equivalent to 5 to 6% oxygen in the flue gas-dry basis) normally encountered in utility power plant operations burning medium or high sulfur coal (coal containing 2.5% sulfur or higher), the level of oxidation is expected to be on the order of 5 to 10% of the sulfur dioxide removed.

2. Absorbent Regeneration

Spent scrubber solution from the absorber recirculation line is bled to this regeneration reactor system where it is reacted with hydrated lime. The reactor system incorporates a novel design developed to produce solids with good settling and filtration characteristics over a broad range of concentrations of sulfite and sulfate in the feed liquor. The reactor train consists of two reactor stages: a short residence time first stage (3-15 minutes) followed by a longer residence time second stage (20-40 minutes). The normal mode of operation is to feed hydrated lime slurry to the first stage reactor only. The process can be operated in conjunction with a lime slaker or can use carbide sludge.

The rate of the lime feed is controlled by the pH of the secondary reactor liquor. The lime neutralizes the bisulfite acidity in the scrubber bleed and further reacts with sodium sulfite and sodium sulfate to produce sodium hydroxide. These reactions precipitate mixed calcium sulfite and sulfate solids, resulting in a slurry containing up to 5 wt % insoluble solids. No recycle of solids is necessary to produce waste material with good properties. The overall reactions are shown below.

$$2NaHSO_3 + Ca(OH)_2 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2\ H_2O\downarrow + 3/2\ H_2O$$

$$Na_2SO_3 + Ca(OH)_2 + 1/2\ H_2O \rightarrow 2NaOH + CaSO_3 \cdot 1/2\ H_2O\downarrow$$

$$Na_2SO_4 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_4\downarrow$$

The CEA/ADL dual alkali process is designed to operate in a relatively "concentrated" active sodium mode (roughly 0.5M active Na^+). In this mode, sulfate removal cannot be affected by precipitation of gypsum ($CaSO_4 \cdot 2H_2O$), since the high sulfite levels prevent soluble calcium concentrations from reaching levels required to exceed the gypsum solubility product; the system is unsaturated with respect to calcium sulfate. However, calcium sulfate is coprecipitated with calcium sulfite in the regeneration reactor. The amount of this sulfate coprecipitated is a function of the ratio of concentrations of sulfate to sulfite in solution and the reactor pH.

At any active sodium concentration above about 0.15M and oxidation rate less than about 25%, the soluble sulfate level in the system will self-adjust to achieve a concentration at which the rate of sulfate coprecipitation as a calcium salt is equal to the rate of sulfate formation by oxidation in the system. The higher the oxidation rate, the higher the steady-state soluble sulfate concentration in the solution. At an active sodium concentration of 0.5M and at a 10% oxidation rate, a soluble sulfate concentration of about 0.5M will produce waste solids containing 10% of the sulfur as calcium sulfate. Under these conditions the waste filter cake would contain 60-65% insoluble solids and could be readily washed to a soluble solids level of 1-2% (dry cake basis).

As the rate of oxidation increases, the sulfate to sulfite ratio in the solution fed to the reactor will increase to keep up with the rate of oxidation. The relationship between the sulfate to sulfite in the regenerated liquor and the amount of calcium sulfate precipitated is plotted in Figure III-2. As shown, the system is capable of keeping up with oxidation rates equivalent to 25% of the SO_2 absorbed without purging sodium sulfate via coprecipitation of calcium sulfate with calcium sulfite. Thus, the solution remains unsaturated with respect to calcium sulfate, thereby circumventing any increase in soluble calcium concentration and any scaling hazards. The sulfate/sulfite ratio can obviously be increased by increasing sulfate concentrations in the liquor and/or by decreasing the sulfite or active sodium concentration. However, it is desirable to keep the active sodium concentration from dropping below about 0.15M, because at this concentration the system becomes "dilute" in active sodium and becomes saturated or supersaturated in calcium sulfate. Under these conditions, soluble calcium concentrations rise to levels of 400-800 ppm and cake properties generally deteriorate.

In practice the system would normally be operated at a total sodium concentration consistent with the highest sustained level of oxidation expected--that is, a sodium concentration high enough that under the worst conditions with regard to oxidation (up to 25-30% oxidation) the active sodium concentration would not fall below about 0.15M Na^+. The liquor composition can

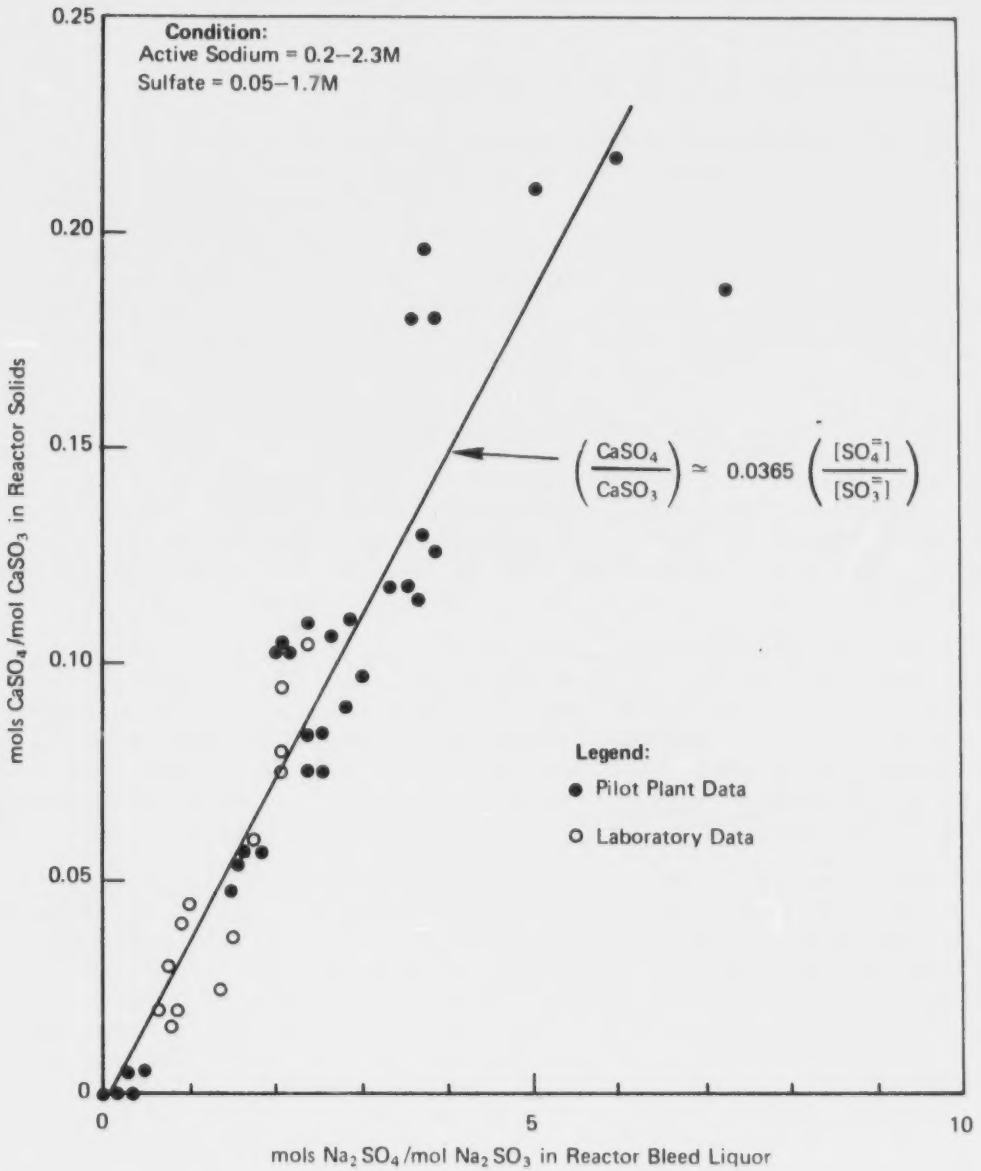

FIGURE III–2 ($CaSO_4/CaSO_3$) RATIO IN REACTOR SOLIDS AS A FUNCTION OF REACTOR LIQUOR COMPOSITION

then vary with changes in oxidation, the sulfate/sulfite ratio adjusting to whatever level is required to keep up with sulfate formation in the system.

3. Solid/Liquid Separation and Solids Dewatering

Slurry from the regeneration reactor system is fed to the center well of the thickener. The thickened slurry from the bottom of the settler is sent to a rotary drum vacuum filter. The slurry is recirculated past the filter in a recycle loop that returns the slurry to the solids zone in the settler. The feed to the filter is drawn as the bleed from this recirculation loop. The filter is equipped with an overflow pipe returning to the solids zone in the thickener to provide protection against inadvertent overflow of the filter hold tank as well as to allow for operation in an overflow mode.

The solid cake is washed on the filter using a series of water spray banks. This wash removes a large fraction (approximately 90%) of the occluded soluble salts from the cake and returns these salts to the system, thereby reducing sodium losses and minimizing sodium carbonate makeup. The total wash rate will usually be set to be a constant percentage of the cake discharge rate. The mixed filtrate and wash liquor from the filter are returned to the thickener.

Makeup sodium carbonate solution is fed to the thickener center well at a rate based upon the lime feed rate to the regeneration reactor system. The sodium makeup rate is tied to the lime feed, as the amount of sodium loss from the system will be a constant percentage of the solids discharged from the filter (for a given wash rate and cake properties). Over a reasonable period of time the solids discharge rate will in turn be in direct proportion to the lime feed.

The sodium carbonate can be fed to the thickener in order to allow easy removal of any $CaCO_3$ precipitated or the sodium carbonate can be fed directly to the scrubber system with no $CaCO_3$ precipitation. The sodium carbonate in the thickener will react with calcium sulfite producing sodium sulfite and calcium carbonate. The calcium carbonate will precipitate in the thickener. This represents a loss of calcium because it is associated with carbonate and not with sulfite or sulfate. On the other hand, the sodium carbonate in the abosrber will react with the sodium bisulfite producing sodium sulfite and carbon dioxide. The carbon dioxide is librated in the absorber. Thus, calcium carbonate is not generated and there is no loss of calcium.

The sodium carbonate is not intended for use as a softener, since soluble calcium concentrations in the regenerated liquor will run less than 100 ppm.

Clear liquor overflow from the thickener is collected in the thickener hold tank which acts as a small surge capacity for the absorbent liquor feed to the scrubber system. Water is also added to this hold tank to

makeup for the difference between total system water losses (evaporation and cake moisture) and total water inputs from other sources (sodium make-up solution, pump seals, lime feed, and cake wash).

B. POLLUTION CONTROL CAPABILITIES - CEA/ADL DUAL ALKALI PROCESS

1. SO_2 Control

The sodium-based dual alkali process, operating in the concentrated active sodium mode, is capable of SO_2 removal efficiencies in excess of 95% over any range of inlet SO_2 concentrations encountered in coal-fired utility boiler applications. In most cases, removal efficiencies approaching 99% can be achieved on a continuous basis. This is accomplished by proper selection and design of the absorber unit and by adjustment of the active sodium/SO_2 stoichiometry in the absorber. Such variation in SO_2 absorption efficiency can be affected without influencing the overall lime stoichiometry or the sodium makeup requirement. These high SO_2 removal efficiencies can be achieved in tray-type absorbers at low L/G ratios, typically in the range of 5-10 gpm/1,000 acfm of saturated gas. The pressure drop across the trays may be minimal, in the range of 4-6 inch WG.

The high SO_2 removal capability of this process, when used in conjunction with a boiler equipped with adequate particulate control, allows the option of removing virtually all of the SO_2 from the flue gas treated in the scrubber and bypassing hot, untreated gas to provide part or all of the reheat while still meeting the overall plant SO_2 emission regulations in the combined treated and untreated flue gas. In such a system the scrubber size can be reduced, since not all flue gas is treated, and the reheat requirements are reduced or eliminated.

2. Particulate Control

Particulate removal can be accommodated in the process by appropriate selection of scrubbers to be used for both the particulate removal and the SO_2 absorption duty. If particulate removal is to be accomplished as part of the overall system than a higher energy particulate removal device, such as a venturi scrubber, may be incorporated in this system to provide for both SO_2 and particulate removal. Particulate removal down to 0.02 grains/scfd or lower can be accomplished using venturi scrubbers at moderate pressure drop of about 15 inch WG.

3. Halogen Control

A major fraction of the chlorides in coal (greater than 90%) is volatilized and appears in the flue gas as HCl. Any aqueous-based scrubbing system would be highly effective in absorption of HCl (and any HF) in the flue gas. As a result, chloride concentrations will build in the closed liquor loop to levels such that the rate at which chloride is discharged from the system in the washed cake will equal the rate at which chloride enters the system with the flue gas. Steady-state levels of chloride in the closed liquor loop of a 20 Mw prototype CEA/ADL dual alkali system rose to as high as 11,000 ppm

(0.05-0.1% Cl in coal) with no apparent effect on the process operation (LaMantia, et al., 1977). Tests of the lime regeneration reaction at ADL have shown that lime utilization and solids properties are unaffected by chloride concentrations as high as 25,000 ppm.

C. WASTE PROPERTIES

The only waste material from the CEA/ADL dual alkali process is the solid waste calcium salt mixture which is washed and contains some residual process liquor. There are no other purges from the system.

This solid material will contain approximately 55-70% insoluble solids and will have excellent handling properties due to precipitation of mix crystals of $CaSO_4/CaSO_3$. The material has a very high angle of repose and is easy to manage in solids handling and transport equipment. The solids are non-thixotropic, drain well, and do not reslurry when exposed to rain. If further chemical treatment is required, these excellent handling properties should prove to be an advantage.

The exact chemical composition of the solids will depend somewhat upon fly ash loading, the chemical composition of the flue gas and fly ash, and the degree of oxidation of sulfite to sulfate encountered in the system. For a high sulfur coal application we would expect the following general chemical composition (dry cake basis):

$$CaSO_3 \cdot 1/2\ H_2O = 80\text{-}85 \text{ wt } \%$$

$$CaSO_4 = 10\text{-}15 \text{ wt } \%$$

$$Na_2SO_4 + Na_2SO_3 + NaCl = 1\text{-}2 \text{ wt } \%$$

$$CaCO_3 + \text{Inerts} = 5\text{-}10 \text{ wt } \%$$

Results of the EPA/ADL dual alkali program indicate that the calcium sulfate is coprecipitated with the calcium sulfite and is not present as gypsum in solids from the concentrated dual alkali mode (LaMantia, et al., 1977). The sodium salts in the washed cake at these low levels of sodium content appear to be occluded in the calcium salt crystals and are difficult to wash or leach out to lower levels. Sodium losses at these levels are negligible as an element of the overall cost of operating the system.

The waste material will also contain chloride (as NaCl) and other volatile constituents such as fluorides, in the coal as sodium salts which are removed in the scrubbing system. The levels of these constituents will build in the closed liquor loop to levels at which the losses in the washed cake become equal, at steady state, to the rate at which the trace constituents are scrubbed from the flue gas. This problem is similar to that encountered in any wet scrubbing system, even recovery systems which employ a "prescrubber loop" to protect the closed loop chemical system.

IV. DUAL ALKALI DEMONSTRATION SYSTEM AT LG&E

This chapter provides the description of the dual alkali system under construction on LG&E Cane Run Unit No. 6. At the Cane Run Plant, lime slurry scrubbing processes have been installed on Units 4 (180 Mw) and 5 (170 Mw).

A. BOILER SYSTEM DESCRIPTION

Cane Run Unit No. 6 consists of a pulverized coal-fired steam generator, built by Combustion Engineering, with a Westinghouse turbine-generator. The unit operates from a minimum of 60 Mw during off-peak hours to a maximum load of 300 Mw during peak hours. The annual average load is equivalent to approximately 180 Mw (about 60% of the gross peak capacity).

In the present system, flue gas from the steam generator passes, in parallel streams, though two Ljungstrom combustion air preheaters. Each air preheater discharges flue gas through separate ducts to an electrostatic precipitator designed for 99.4% particulate removal efficiency. From the precipitator the gases enter two induced draft fans (each handling 50% of the total gas) and the two gas streams are then combined before entering the stack.

The sulfur dioxide system will be installed between the existing induced draft fans and the chimney. Hot flue gas will be drawn from the existing ducting at the outlet of the induced draft fans and fed to the dual alkali scrubber system through two booster fans. The scrubber gas will be reheated and then returned to the existing entrance to the stack. Appropriate dampers will be provided to allow bypass of gas around the scrubber system using the existing duct.

Coal for Unit No. 6 is received from a number of sources. A dry ultimate analysis given in Table IV-1 being typical of the coal fired. The average sulfur content on a dry basis is 4.8% and varies from 3.5% to 6.3%. The average chloride content of the coal is 0.04% and varies from 0.03% to 0.06%. The average 4.8% sulfur content and 11,000 Btu/lb will result in an average SO_2 emission of about seven times of those allowed by the present Federal New Source Performance Standards (1.2 lbs of SO_2/MM Btu).

The flue gas conditions at the exit of the existing induced draft fans are given in Table IV-2. The gas flow rate through each fan is controlled to be equal. The gas flow rate shown in Table IV-2 represents the total gas flow.

B. DESIGN BASIS FOR THE DUAL ALKALI SYSTEM

The design basis for the dual alkali system is summarized in Table IV-3. Design conditions correspond to coal containing 5% sulfur and 0.04% chloride, and having a heating value of 11,000 Btu/lb. The design gas capacity of 3,372,000 lbs flue gas/hr is equivalent to the boiler peak load capacity of

TABLE IV-1

ULTIMATE COAL ANALYSIS

(Dry Basis)

	Typical Analysis, %	Range, %
Carbon	67.15	64.0-70.0
Hydrogen	4.72	4.3-5.25
Nitrogen	1.28	0.6-1.5
Chlorine	0.04	0.03-0.06
Sulfur	4.81	3.5-6.33
Ash	17.06	15.5-24.5
Oxygen	4.94	3.8-6.2
Moisture	8.95	8.0-10.75
Heat Content, Btu/lb	11,000	9,500-12,400 Maximum 10,400-11,900 Normal

TABLE IV-2

FLUE GAS CONDITIONS AT THE INLET OF THE DUAL ALKALI SYSTEM

Normal Operating Temperature	300°F
Maximum Gas Temperature for Periods up to 5 Mins.	600°F
Normal pressure at I.D. Fan Outlet	-1" to +2" WG
Boiler Excess Air	25% Maximum 35%
Air Heater Leakage	10%
Flue Gas Density at Sea Level @ 70°F	0.078 lb/ft^3
Total Pressure at Stack Entrance	+2" WG
Boiler Load Points (lbs/hr flue gas):	
Design	3,372,000
Boiler Maximum Continuous Rating	3,003,000
Control Load	1,440,000
Minimum Normal Operating Load	658,000

TABLE IV-3

DESIGN BASIS

Coal (Dry Basis):	
Sulfur	5.0% S
Chloride	0.04% Cl
Heat Content	11,000 Btu/lb
Inlet Gas:	
Flow Rate (Volumetric) (Weight)	1,065,000 acfm 3,372,000 lb/hr
Temperature	300°F
SO_2	3471 ppm
O_2	5.7%
Particulate	$\leq$0.10 lb/10^6 Btu
Outlet Gas:	
SO_2	$\leq$200 ppm ($\sim$0.45 lb/10^6 Btu)
Particulate	$\leq$0.10 lb/10^6 Btu

300 Mw. The dual alkali system will meet all applicable federal, state, and local pollution control and safety regulations. The maximum SO_2 concentration in the clean gas will be 200 ppm (for coal containing up to 5% sulfur), well below requirements of the Federal NSPS.

At present, Cane Run Unit No. 6 is operating with all applicable federal, state, and local operating permits for air, water (NPDES), OSHA, etc. Construction permits have been issued by the Air Pollution Control District of Jefferson County District (APCDJC) for the construction of an SO_2 removal system for Cane Run Unit No. 6. The permit complies with the two enforcement orders with the EPA (November 5, 1975) and the APCDJC (December 10, 1975).

Disposal of wastes from the dual alkali system will meet all applicable federal, state, and local solid waste disposal regulations in effect at present. None of the wastes produced by the control system will be discharged to or allowed to enter any naturally occurring surface water. A long-range disposal plan is under development at this time. Such a plan will be available by July 1, 1977 in accordance with present agreement under the NPDES Permit Regulations. A preliminary disposal plan is presented as part of this manual.

The existing 518 foot stack will be the only source of gaseous emissions from the system. A flue gas bypass will be provided to allow untreated boiler flue gas to enter the stack, bypassing the control system scrubber. The scrubber system is designed to be completely isolated from the flue gas during periods in which the bypass is open, to allow safe entry into the scrubber system for maintenance and inspection while the boiler continues normal operation. Also, each absorber can be isolated independently and maintenance can be provided to the absorber while the other absorber is in operation. The duct dampers are designed such that with the dual alkali system in operation and the bypass closed, no more than 1.0% of the total flue gas will leak through the bypass system into the stack.

The dual alkali system will be equipped with sufficient instruments, in addition to those required to operate the process, to permit accurate measurements of the appropriate streams required to calculate material and energy balances. In particular, instrumentation will be provided to permit continuous monitoring of SO_2 concentrations in the flue gas entering and leaving the control system as well as the measurement of the quantities of all chemicals and water entering the system.

C. GUARANTEES

1. Sulfur Dioxide Emission

The system shall provide such control that emissions from the stack shall be no greater than 200 ppm SO_2 dry basis without additional air dilution when burning coal containing less than 5% sulfur. When burning coal containing 5% sulfur or greater, the system shall be designed to remove at least 95% of the sulfur dioxide in the inlet flue gas.

2. Particulate Emission

In addition to meeting applicable regulations, the system shall also be designed to meet Federal New Source Performance Standards for emissions of particulates under all conditions of boiler operation. The dual alkali system will not add any particulate matter to the emissions of particulate matter that is received by the system from the LG&E Cane Run Unit No. 6 electrostatic precipitator.

3. Lime Consumption

The consumption of lime in the system will not exceed 1.05 moles CaO per mole of SO_2 removed from the flue gas.

4. Sodium Carbonate Makeup

Soda ash makeup will not exceed 0.045 moles of Na_2CO_3 per mole of SO_2 removed from the flue gas, provided that the chloride content of the coal burned averages 0.06% or less. If the average chloride content of the coal is above 0.06%, then additional sodium carbonate consumption will be allowed at the rate of 1/2 mole Na_2CO_3 for each mole of chloride in the flue gas resulting from chloride in excess of 0.06% in the coal. The seller as part of the guarantee shall perform the necessary research and design to reduce the makeup requirements of Na_2CO_3 from the guarantee point to a level approaching minimal makeup.

5. Power Consumption

At the peak operating rate (300 Mw) the system will consume a maximum of 1.1% of the power generated by the unit.

6. Waste Solids Properties

The waste produced by the vacuum filter will contain a minimum of 55% insoluble solids.

7. SO_2 System Availability

The system will have an availability (as defined by the Edison Electric Institute for power plant equipment) of at least 90% for one year. Thus, the system will be available for operation at least 90% of the calendar time.

D. GENERAL DESCRIPTION OF THE DUAL ALKALI SYSTEM

The dual alkali system will use carbide lime and soda ash as raw chemicals. Carbide lime is a byproduct of acetylene production and is available locally at a significantly lower price compared to commercial lime. The carbide lime will be barged to the plant in the form of slurry containing 30% solids. The slurry is unloaded at the Cane Run Plant and stored in a tank. The specifications of carbide lime are given in Table IV-4.

TABLE IV-4

CARBIDE LIME SPECIFICATIONS

	Carbide Lime Slurry[a]
Calcium hydroxide $Ca(OH)_2$	92.50
Available calcium oxide CaO	70.01
Calcium carbonate $CaCO_3$	1.85
Silica SiO_2	1.50
Iron and alumina oxides R_2O_3	1.60
Magnesium oxides MgO	0.07
Sulfur	0.15
Phosphorus	0.01
Free carbon	0.25
Free water	--
Not analyzed	2.07

[a]Available as slurry containing 30% insoluble solids.

Source: Airco catalog (1969).

The lime is specified to have a $Ca(OH)_2$ content of 92.5% and a particle size distribution equivalent to 90% through a 325 mesh screen.

As discussed previously, the dual alkali system can be divided into three major sections: absorber section; reactor section; and solids/liquid separation section. The system design is modular in nature with spare capacity provided both as excess capacity within modules as well as spare modules and equipment where appropriate. Also, spare instrumentation has been provided for critical control operations.

In addition to the dual alkali system, a sludge disposal system will be installed at the Cane Run Plant. The details of the sludge disposal are given in the later part of this chapter.

The process flowsheet for the system is shown in CEA drawings[a] 040044-1-1 and 040044-1-2, and the plot plan is shown in CEA drawing[a] 040044-2-0 (not included here). The flue gas from the Cane Run Unit No. 6 boiler passes through two existing electrostatic precipitators and two existing induced draft fans before entering the dual alkali system.

1. Absorption Section

The flue gas from the existing induced draft fan is forced by a booster fan into an absorber. There are two absorber modules and each module is equipped with a booster fan. A common duct connects the two inlet ducts to the booster fans.

The hot flue gas is cooled by sprays of absorber solution directed at the underside of the bottom tray. These sprays keep the underside of the tray and the bottom of the absorber free of buildup of fly ash solids. The cooled gas then passes through a set of two trays, where SO_2 is removed, then through a chevron type mist eliminator. Prior to entering the stack, the saturated gas from the demister is heated 50F°, to 175°F, by hot flue gas from a reheater fired with No. 2 oil.

The absorber design is based upon full scale and pilot experience with similar sodium absorption systems. The design is conservatively based upon 9.0 fps gas velocity, a rate consistent with good mass transfer, low pressure drop, and minimum entrainment. Each absorber is sized to handle 60% of the design gas flow rate and the system can be turned down to 20% of the design flow rate. At levels less than 60% of the design capacity the system can be operated with one absorber module (by using common duct and shutting down one absorber module) or with two absorber modules.

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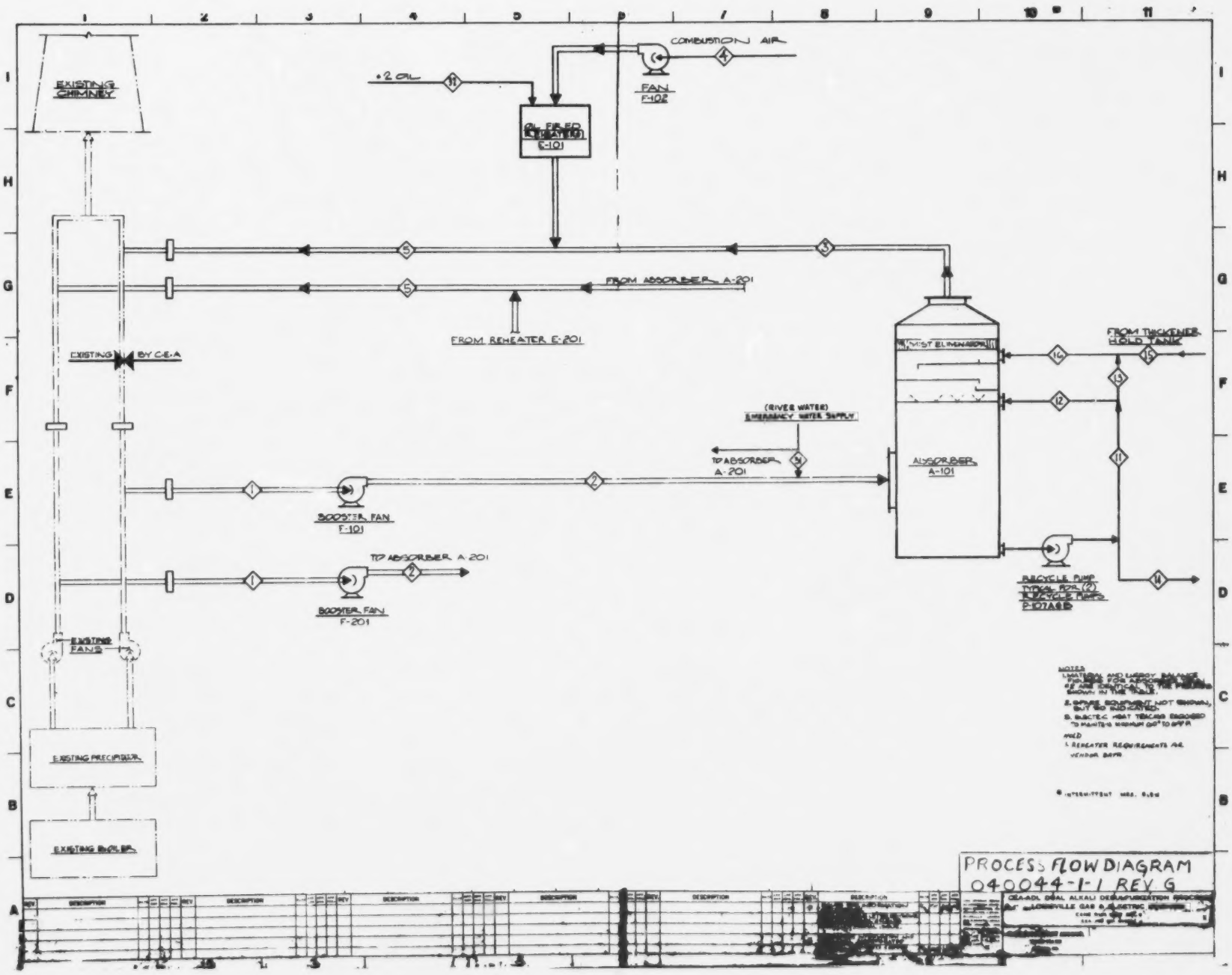
[a] A set of large size CEA drawings for this system may be obtained from Environmental Protection Agency, Research Triangle Park, North Carolina. Because of too many details, these drawings cannot be reduced; therefore, most of them are not included here.


EXISTING CHIMNEY
COMBUSTION AIR
FAN F-102
#2 OIL
OIL FIRED REHEATER E-101
FROM ABSORBER A-201
FROM REHEATER E-201
EXISTING BY C-E-A
MIST ELIMINATOR
FROM THICKENER HOLD TANK
(RIVER WATER) EMERGENCY WATER SUPPLY
TO ABSORBER A-201
ABSORBER A-101
BOOSTER FAN F-101
TO ABSORBER A-201
BOOSTER FAN F-201
RECYCLE PUMP TYPICAL FOR (2) RECYCLE PUMPS P-107A&B
EXISTING FANS
EXISTING PRECIPITATOR
EXISTING BYPASS
NOTES
1. MATERIAL AND ENERGY BALANCE FOR ABSORBER ONLY. OTHER PARTS OF THE PROCESS ARE SHOWN IN THE TABLE.
2. BYPASS EQUIPMENT NOT SHOWN, BUY AND INSTALL.
3. ELECTRIC HEAT TRACING PROVIDED TO MAINTAIN MINIMUM 60° TO DUCT
4. REHEATER REQUIREMENTS AS VENDOR DATA
PROCESS FLOW DIAGRAM
040044-1-1 REV G

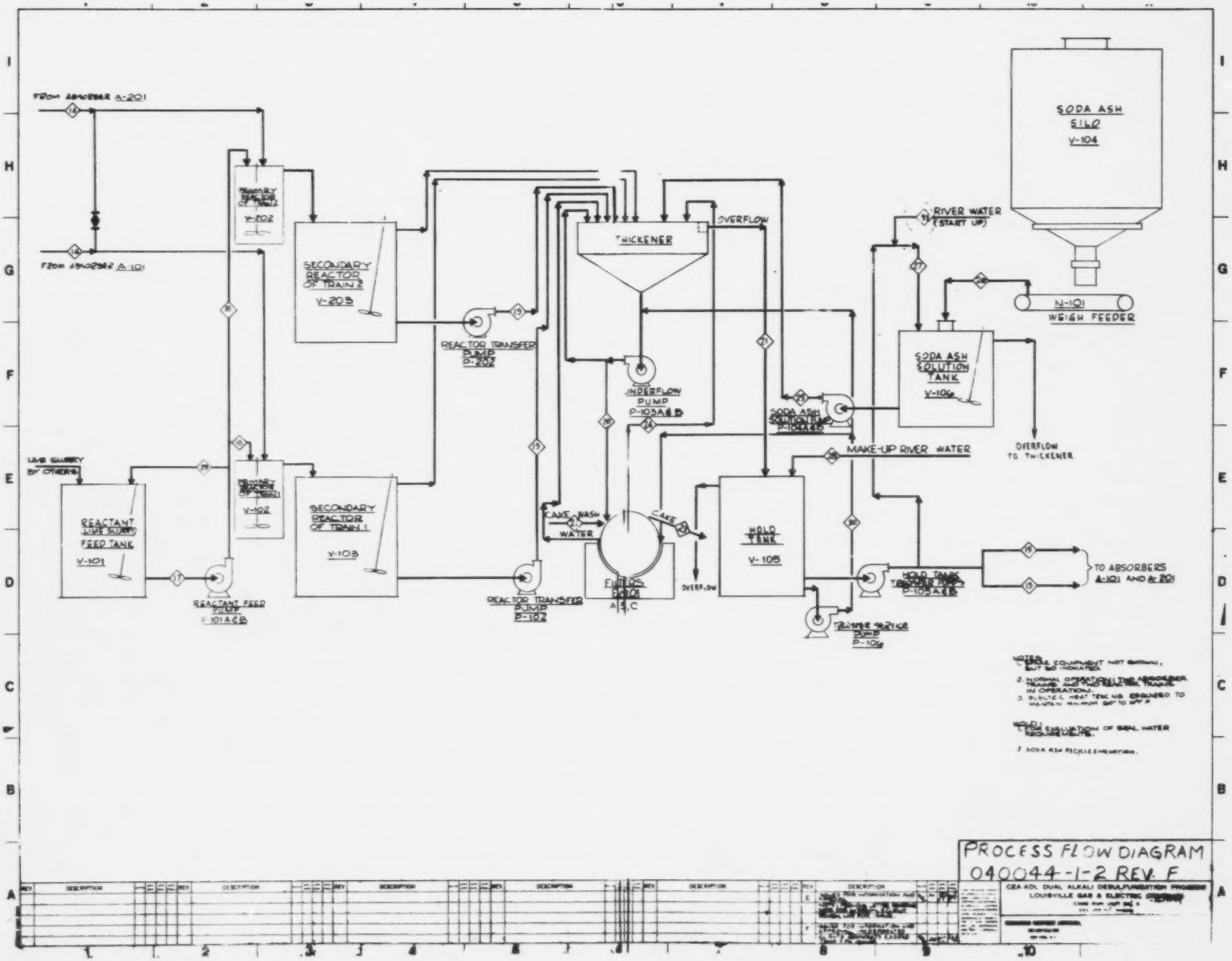
Stream No.	1	2	3	4	5
Volume, ACFM	532,907	525,880	436,516	11,000	487,215
Temperature, ^{o}F	300	303	126	60	176
Pressure, inches WG	+2.0	+11.5	+3.5	AMB	+2.0
Dry Gas, #/min	26,699	26,699	26,515		
H_2O Vapor, #/min	1,316	1,316	2,474		
Total Gas, #/min	28,015	28,015	28,989	850	27,839
SO_2, #/min	194.95	194.95	11.25		11.25
SO_2, PPM (Dry Vol)	3,471	3,471	200		200
Particulates, #/min	2.48	2.48	2.48		2.48

	11	12	13	14	15	16	31	32
H_2O, #/min	19,933.5	12,919.5	7,014.0	13,642.8	14,828.1	21,867.0	4,582.0	
Na_2CO_3, #/min					6.8			
NaOH, #/min					50.0			
Na_2SO_3, #/min	174.9	113.4	61.5	119.7	347.2	582.0		
$NaHSO_3$, #/min	666.9	432.2	234.7	456.4		95.0		
Na_2SO_4, #/min	1,542.9	1,000.0	542.9	1,056.0	1,016.3	1,559.2		
$Ca(OH)_2$, #/min								
$CaSO_3 \cdot \frac{1}{2}H_2O$, #/min								
$CaSO_4 \cdot \frac{1}{2}H_2O$, #/min								
$CaCO_3$, #/min								
Inerts, #/min								
NaCL, #/min	445.8	288.9	156.9	305.1	303.6	460.5		
Total, #/min	22,764	14,754.0	8,010.0	15,580.0	16,552.0	24,563.7	4,582.0	
% Solids, #/min	0	0	0	0	0	0	0	
Flow, GPM	2,481	1,608.0	873.0	1,698.0	1,837.0	27,101.0	550.0*	3.0
pH	4-8	4-8	4-8	4-8	7-13	6-11		
Temp, ^{o}F	120-140	120-140	120-140	120-140	120-140	120-140		AMB.

*Intermittent Max. Flow

Combustion Equipment Assoc.

Material Balance
for
Drawing #040044-1-1 Rev. G



NOTES:
1. [illegible] EQUIPMENT NOT SHOWN, BUT SO INDICATED.
2. NORMAL OPERATION WITH THE ABSORBER TRAINS AND [illegible] IN OPERATION.
3. [illegible] HEAT TRACING TO BE ENERGIZED TO MAINTAIN MINIMUM [illegible] TO [illegible]

WORK:
1. FOR EVALUATION OF [illegible] WATER REQUIREMENTS.
2. SODA ASH RECYCLE EVALUATION.

PROCESS FLOW DIAGRAM
040044-1-2 REV. F

CEA ADL DUAL ALKALI DESULFURIZATION PROCESS
LOUISVILLE GAS & ELECTRIC [illegible]

REV	DESCRIPTION
E	[illegible]
F	[illegible]

Stream No.	14	15	17	18	19	20	21	22	23	24	25	26	27	28	29	30	33
H_2O, #/min	13,642.8	14,828.1	2,150.8	537.7	14,233.0	2,108.5	29,182.8	756.0	442.1	2,421.4	403.9		403.9	878.0	1,075.4		
Na_2CO_3, #/min		6.8					13.9				13.9	13.7	0.2				
NaOH, #/min		50.0			50.2	7.3	101.3		0.50	6.8	1.4		1.4				
Na_2SO_3, #/min	119.7	347.2			349.4	51.0	704.9		3.4	47.5	9.6		9.6				
$NaHSO_3$, #/min	456.4																
Na_2SO_4, #/min	1,056.0	1,016.3			1,019.8	148.7	2,057.6		10.0	138.8	27.9		27.9				
$Ca(OH)_2$, #/min			852.0	213.0											426.0		
$CaSO_3 \cdot \frac{1}{2}H_2O$, #/min					331.0	662.0			662.0								
$CaSO_4 \cdot \frac{1}{2}H_2O$, #/min					39.4	78.8			78.8								
$CaCO_3$, #/min					5.8	11.6			11.6								
Inerts, #/min			69.2	17.3	17.3	34.6			34.6						34.6		
NaCl, #/min	305.1	303.6			305.1	44.5	615.5		3.00	41.5	8.3		8.3				
Total, #/min	15,580.0	16,552.0	3,072.0	768.0	16,351.0	3,147.0	32,676.0	756.0	1,246.0	2,656.0	465.0	13.7	451.3	878.0	1,536.0		
% Solids	0	0	30.0	30.0	2.4	25.0	0	0	63.1	0	0				30.0		
Flow, GPM	1,698.0	1,837.0	305.2	76.3	1,785.0	302.0	3,618.0	91.0		299.0	50.0		50.0	105.0*	152.6	250.0**	135.0 max.
pH			10-14	7.0	6-13	7-13	7-13			7-13	8-14		7-13			7-13	
Temp, °F					120-140	120-140	120-140	AMB.		120-140	120-140		120-140	AMB.		120-140	

*Includes water entering process from pump seals
**Intermittent Maximum

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Combustion Equipment Assoc.

Material Balance
for
Drawing 040044-1-2 Rev F

For control of tray feed liquor pH, regenerated scrubbing liquor from thickener hold tank is mixed in line with absorber recycle liquor for each unit; the mixture is then fed to the top tray in each absorber. The absorber recycle liquor is used in the spray section below the trays. A bleedstream of the absorber recycle liquor is withdrawn and sent to the reactor system for regeneration. The bleed rate is controlled by the liquid level in the absorber. The feed forward of regenerated liquor from the thickener hold tank to the scrubber trays corresponds to an L/G of 4.0 gals./1,000 acf (saturated) at design conditions. The total recirculation rate for each absorber (sprays plus trays) corresponds to an L/G of 5.7 gals./1,000 acf.

The overall gas pressure drop through the dual alkali system is less than 9.5 inches WG and includes a pressure drop of 4-6 inches WG across the absorber trays.

On the gas side, appropriate interlocks are provided in the bypass and FGD system inlet dampers to enable bringing scrubbers on- or off-line without interruption of the boiler operation. The speed of each of the booster fans is controlled to maintain the desired pressure level at the outlet of the existing boiler induced draft fans (with load input from the boiler control system). Pressures and temperatures are measured at appropriate points throughout the gas flow and the scrubber system. SO_2 is monitored by continuous analyzers at the inlet and outlet of the absorber. Provision is made for emergency water supply for automatic quenching, if recycle pumps fail.

2. Reaction Section

The spent liquor is introduced to the primary reactor of the two-stage reactor system along with slurried carbide lime from the carbide lime storage tank for the dual alkali system (day tank). The primary reactor has a nominal liquor holdup time of 4.5 minutes at design flow. The primary reactor overflows by gravity into the secondary reactor, which has a nominal liquor holdup of 40 minutes at design flow, where the reaction between lime and sodium salts is completed. The reaction product is a slurry containing 2-4% insoluble calcium salts and the regenerated sodium salt solution. The reactor liquor is pumped to the solid/liquid separation section. The pumping rate is controlled by the liquid level in the reactor.

There are two reactor trains in the system and each train consists of a primary reactor, a secondary reactor, and a reactor pump. Normally, the number of reactor trains in use is the same as the number of operating scrubber modules, with each reactor train regenerating the spent liquor from the corresponding scrubber module. Both the reactor trains are sized identically; each module can be operated on liquor from the corresponding scrubber or liquor from both the scrubbers. For short term, a reactor train can handle the total liquor from both the absorbers operating at design conditions. Thus, maintenance can be provided to one reactor train while the system is operated with the other.

The carbide lime slurry used from regeneration of the spent absorber solution is received as a 30% solids slurry and stored in a large tank for use in the three FGD systems at Cane Run. From this main storage the slurry is pumped to a day tank for each system. Slurry from the dual alkali system day tank is pumped to the primary reactors as required. The slurry feed rate to the primary reactor is controlled by the pH of the liquor in the corresponding secondary reactor. For example, the slurry feed rate may be controlled to maintain a pH of 12.0 in the secondary reactor.

3. Solids/Liquid Separation

The reactor effluent streams are fed to the thickener. Clarified liquor overflows to the thickener hold tank from which the regenerated solution is pumped to the absorbers as required. As discussed before, the feed rate to the absorbers is determined by the pH of the absorber liquor. The total volume in the system is maintained by controlling the liquid level in the thickener hold tank using process makeup water.

The thickener underflow slurry controlled at about 25% solids is pumped to the filter system where solids separation is completed. The filter cake is washed to recover the sodium salts in the liquor. Combined filtrate and wash water is returned to the thickener.

There are three filters, each rated to handle 50% of the total solids produced at the design conditions. Each filter can be operated independently. For optimum performance (to obtain cake containing high dry solids and low soluble salts) it is desirable to operate the filters at fixed conditions (constant drum speed, submergence, wash ratio, etc). Therefore, the cake rate is controlled by changing the number of filters in operation. The number of filters in operation is determined by the quantity of solids accumulated in the thickener, which is reflected in the solids concentration in the underflow slurry. The density of the underflow slurry is measured and thickener hold tank liquor is added as required to maintain the percent solids in the underflow slurry to about 25%. The number of filters in operation is changed if the concentration of solids in the underflow slurry cannot be controlled using the dilution liquor.

Soda ash is added to the system to make up for the sodium salts lost in the waste filter cake. Dry, dense soda ash is received at the plant and stored in the soda ash silo from which it is continuously weighed and fed to the soda ash solution tank. The soda ash feed rate is generally controlled in proportion to the lime slurry feed rate. Over the long run the lime slurry feed is proportional to the filter cake produced, and thus, the sodium losses in the cake. The soda ash solution is prepared using thickener hold tank liquor, which is fed to the soda ash solution tank at a constant 50 gpm. Soda ash solution is pumped to the thickener.

4. Spills and Leakages

Filter cake is the only product of the dual alkali process. The system is operated in a closed loop, and there is no other liquid or solid purge from the system. In practice, some liquor can be lost, though, through pump seals and during flushing of pipes and pumps after individual pumps are shut down or after the system is shut down. The concentration of the liquor is generally low compared to the concentration in the process liquor because of dilution with seal or flush water. However, to prevent water pollution and to reduce sodium loss from the system, this solution is collected in sump tanks and is returned to the thickener.

5. Waste Disposal

A long-range plan for the disposal of the dual alkali filter cake is now being developed as a part of an overall disposal plan for all FGD sludges produced at the Cane Run Station. The plan, in its preliminary form, will be submitted to the Jefferson County Air Pollution Control Board in July 1977.

As currently conceived, the disposal operation will involve mixing dual alkali filter cake with thickener underflow from the direct lime scrubbing systems on Cane Run Unit Nos. 4 and 5, producing a material of approximately 40% solids. The combined wastes will then be transported to an onsite disposal pond. Prior to being placed in the pond, the combined sludge will be admixed with dry fly ash collected from the electrostatic precipitators and separately conveyed to the disposal area. Additives, such as lime, may also be used to promote hardening of the material.

Studies are now underway at the University of Louisville to develop adequate physical properties data to allow for the design of appropriate handling, mixing, and transport facilities. It is hoped that the dual alkali filter cake and direct lime thickener underflows can be handled as a thick slurry capable of being piped to the disposal area.

Since this combined disposal system is still in planning, the necessary handling and transport equipment will not be operational at the time the dual alkali system starts up. Thus, in the interim between the startup of the dual alkali system and installation of the equipment for the combined waste disposal operation, the filter cake will be conveyed to a feed hopper where it will be loaded into trucks for transport to the disposal pond.

V. MATERIAL AND ENERGY BALANCES

Cane Run Unit No. 6 has a rated capacity of 280 Mw and a peak load capacity of 300 Mw. The system is designed to handle the peak load, and the material and energy balances have been calculated for the design conditions. The basis for the material and energy balances is given in Table V-1. The design coal contains 5% sulfur and 0.04% chloride. All estimates are based on 94% of the sulfur in the coal appearing in the flue gas. Process flowsheets with detailed material balances for the system are given in CEA Drawings 040044-1-1 and 040044-1-2 in Chapter IV.

A. MATERIAL BALANCES

1. Overall Balance

The overall balance for the dual alkali system is shown in Table V-2. Filter cake is the only waste product from the system. The weight of the filter cake (1,246 lbs/min.) is roughly equal to the weight of the SO_2 and chloride absorbed (370 lbs/min.), plus the weight of net bound and unbound water in the cake (394 lbs/min.), plus lime (461 lbs/min. as $Ca(OH)_2$ and inerts), and soda ash (14 lbs/min.). The small difference is due to oxidation in the system and evolution of carbon dioxide from the soda ash.

The total gas flow rate at the inlet of the system is 1,065,000 acfm at a temperature of 300°F. Gas is supplied from two existing induced draft fans located downstream of the existing precipitators. The gas flow rate for each fan is 50% of the total gas flow. Two absorber trains are used in the dual alkali system; each train has its own booster fan to force the gas through the absorber. The total saturated gas flow rate from both absorbers is 873,000 acfm at 126°F. Combustion gas from the oil-fired reheater is mixed with saturated gases to reheat the exhaust gases 50°F. The total flow rate of reheated gas from the two trains is 974,000 acfm at 176°F.

The total SO_2 in the combined inlet gas to the two scrubbers at design conditions is 390 lbs/min., which corresponds to an SO_2 concentration in the absorber inlet gas of 3,471 ppm on a dry volumetric basis. The SO_2 concentration in the absorber outlet gas is 200 ppm. This is equivalent to a total of 22.5 lbs/min. of SO_2 in the outlet gases, which represents a design removal efficiency of 94.2%.

At design load, regenerated scrubbing liquor is fed to each absorber at a flow rate of 3,675 gpm, which is equivalent to about 4.2 gals/1,000 acf of saturated gas. The total design flow rate to the top trays in the two absorbers including recycle of absorber bottoms is 5,400 gpm, equal to about 6.2 gal/1,000 scf of saturated gas. Absorber bottoms are also recirculated to the spray system in each absorber. The flow rate to the sprays in the two absorbers is 3,200 gpm, equivalent to 3.7 gals/1,000 acf of saturated gas. The spent absorber stream from the two absorbers at the rate of 3,400 gpm is sent to the two reactor trains. The difference between the feed forward to the two absorbers and the bleed from the absorbers

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TABLE V-1

BASIS FOR MATERIAL AND ENERGY BALANCES

Coal (dry basis):

Sulfur	5.0% S
Chloride	0.04% Cl
Heating value	11,000 Btu/lb

Inlet Gas:

Flow rate	1,065,000 ACFM
Temperature	300°F
SO_2	3471 ppm (dry basis)
O_2	5.7 vol. %
Particulate	$\leq$0.1 lb/10^6 Btu

Outlet Gas:

SO_2	200 ppm (dry basis) ($\sim$0.45 lb/10^6 Btu)
Particulate	$\leq$0.10 lb/10^6 Btu

Absorber Feed Concentration:

$NaOH + 2[Na_2CO_3]$	0.09 M
Na_2SO_3	0.18 M

Oxidation and Sodium Makeup Rates:

Oxidation	10% ΔSO_2 (molar basis)
Na_2CO_3	$\leq$4.5% ΔSO_2 (molar basis)

Calcium Feed Stoichiometry:

Solids in slurry	30%
Available $Ca(OH)_2$	92.5% solids
$Ca(OH)_2$ efficiency	98%

Waste Solids:

Wash ratio	2.0 displacement washes
Insoluble solids	$\geq$55 wt %

TABLE V-2

OVERALL MATERIAL BALANCE FOR THE DUAL ALKALI SYSTEM

SO_2	- inlet flue gas	390 lbs/min
	- outlet flue gas	22 lbs/min
	- cake	368 lbs/min
Water	- input to process[a]	325 gpm
	- evaporation in scrubber	278 gpm
	- cake	47 gpm
Lime		460 lbs/min
Makeup soda ash		13.7 lbs/min
Fuel oil	- (to reheat wet gases)	6 gpm
Cake		1,246 lbs/min

Basis: 5% S in coal
94% SO_2 removal
300 Mw boiler load

[a] Includes all water inputs (filter cake wash water, seal water, water in carbide lime slurry, and makeup water).

is 278 gpm, equal to the design rate of evaporation in the two absorbers. This evaporation rate is roughly 1 gpm/megawatt.

As discussed before, at excess air levels equivalent to 5-6% O_2 in the boiler flue gas encountered in utility boiler operations burning coal containing higher than 2.5% sulfur, the level of oxidation is expected to be on the order of 5-10% of the sulfur dioxide removal. As shown in Table V-1, the oxidation in the system is conservatively estimated at 10% of the SO_2 removed. The sulfite and sulfate concentrations indicated in the table are consistent with the level required to precipitate sulfate as a calcium salt in balance with 10% oxidation rate (taking into account losses of Na_2SO_4 in the washed cake). The total active sodium concentration in the absorber feed is set at 0.45 M ($[Na_2SO_3]$ = 0.18 M, [NaOH] = 0.08 M, and $[Na_2CO_3]$ = 0.01 M), which results in a Na_2SO_4 concentration of 0.47 M.

The NaCl concentration in the absorber liquor is equivalent to 0.34 M and represents a steady-state level such that the rate at which NaCl is lost in the washed cake is equivalent to the chloride absorbed from the gas: 1.8 lbs/min.

The feed forward rate of the regenerated liquor from the thickener hold tank to the absorber is controlled by the pH of the absorber bleed. At design conditions the absorber bleed pH is assumed to be 6. At this pH the absorber bleed will contain 0.31 M $NaHSO_3$ and 0.07 M Na_2SO_3. The Na_2SO_4 concentration in the bleed is increased to 0.52 M. The changes in the concentrations in the outlet and the inlet streams reflect 10% oxidation, absorption of 368 lbs/min. of SO_2, and changes in the stream volumetric flow rate due to the evaporation of water in the absorbers.

The design feed rate to the two primary reactors is about 3,400 gpm of spent absorber liquor (1,700 gpm to each reactor). Lime slurry containing 30% solids is also fed to the reactors at a rate of 152 gpm. The calcium hydroxide available in the solids is 92.5%. Thus, the total feed of available calcium hydroxide to the two primary reactors is 426 lbs/min. This rate is equivalent to 1.0 moles of calcium hydroxide per mole of SO_2 removed in the absorbers.

The liquor from the primary reactor overflows into the secondary reactor where the reaction is completed.

The composition of the liquor in the secondary reactor is dependent on the degree of regeneration. For design purposes the secondary reactor is assumed to operate at a pH of 12. At this pH the composition of the secondary reactor liquor is: [NaOH] = 0.085 M, $[Na_2SO_3]$ = 0.19 M, and $[Na_2SO_4]$ = 0.48 M. The soluble calcium concentration in the liquor is less than 100 ppm. The liquor also contains calcium sulfite and calcium sulfate solids. The solids concentration in the slurry is 2.3%. The solids contain 1-2% unreacted lime, 5% inerts (from the carbide lime slurry), about 10% calcium sulfate, and about 84% calcium sulfite (see cake composition in the following paragraphs).

The liquor from the two secondary reactors is pumped at a combined rate of 3,570 gpm to the thickener center well where the solids are allowed to settle. The underflow slurry is recycled around the thickener and a bleed from this recirculation loop is sent to the filters. The concentration of solids in the underflow slurry from the thickener is controlled at about 25% solids.

The filters are operated on overflow with the slurry level in the filter tubs controlled by the position of the overflow weir. At design load, the total slurry feed to the filters is 300 gpm and contains approximately 3,150 lbs of insoluble solids/hr. (25% insoluble solids). About 60% of the total slurry fed to the filters is returned in the overflow to the thickener. The solids are filtered, forming a cake which is washed with water to remove sodium salts in the liquor entrained in the cake. The design wash rate is 90 gpm, which corresponds to a wash ratio of about 2.0 (gals. wash water/gal. entrained liquor). The combined filtrate (wash water and recovered liquor) is returned to the thickener. The washed cake is discharged at a rate of 1,246 lbs/min. from the filter drum, and the composition of the washed filter cake is:

Insoluble salts	- 63% of cake weight
$CaSO_3 \cdot 1/2\ H_2O$	- 84% of insoluble salts
$CaSO_4 \cdot 1/2\ H_2O$	- 10% of insoluble salts
Inerts and unreacted lime	- 6% of insoluble salts
Soluble salts	- 2.2% of insoluble salts
NaCl	- 0.9% of insoluble salts
Other Na salts	- 1.3% of insoluble salts

The rate of soda ash makeup required to replace the sodium value lost in the cake is 13.7 lbs/min., equivalent to 2.3% of the SO_2 absorbed on a molar basis. Clarified, regenerated liquor overflows the thickener at the rate of about 3,600 gpm and is stored in the thickener hold tank from which it is pumped to the absorbers. Makeup water is also added to the thickener hold tank to maintain the overall system water balance. The quantity of makeup water added varies according to the boiler operation and meteorological conditions. The overall system water balance is discussed in the following section.

2. Water Balance

a. Steady-State Balance

In order to operate the dual alkali system in a closed loop over a full range of conditions, it is necessary to have a flexible water balance. The principal factors affecting the water balance are boiler load, sulfur content of the coal, and the degree of cake washing. Table V-3 lists the inputs and outputs for both contact and noncontact water used in the system and shows the steady-state system water balance for operation on 3.8% sulfur coal (the expected average sulfur content). The water balance is given for three different boiler loads--20%, 60%, and 100% of the peak 300 megawatt capacity.

TABLE V-3

WATER BALANCE AT EXPECTED OPERATING CONDITIONS
(All Flow Rates in gpm)

Basis: 3.8% sulfur coal
55% insoluble solids in the cake
2.0 wash ratio

	Load (% of 300 Mw)		
	20%	60%	100%
Process Water Inputs:			
Lime Slurry-Free Water	20.3	62.6	104.3
-Chemically Combined Water	2.0	6.0	10.0
Pump Seal Water	20.3	21.0	21.6
Instrument Purge Water	0.5	0.5	0.5
Filter Cake Wash Water	24.1	72.2	120.3
Rain Water[a]	0.6	0.6	0.6
Total	68.4	162.9	257.3
Noncontact Water Inputs:			
Fan Bearing Cooling Water	175	175	175
Vacuum Pump Seal Water	8	24.2	40.4
Total	183	199.2	215.4
Process Water Outputs:			
Evaporation to the Flue Gas	55.6	166.9	278.2
Cake-Free Water	12.0	36.1	60.1
-Chemically Combined Water	0.5	1.6	2.6
Evaporation to Atmosphere[a]	32.0	32.0	32.0
Total	100.1	236.6	372.9
Noncontact Water Outputs:			
Fan Bearing Cooling Water	175	175	175
Vacuum Pump Seal Water	8	24.2	40.4
Total	183	199.2	215.4
Net Water Added to the System to Close Process Water Balance	31.7	73.7	115.6

[a] Annual average (see Figure V-5 for evaporation from open vessels).

Note: Seal water is assumed to be 20 gpm.

The block diagram in Figure V-1 shows the general breakdown of the water balance by each process section at 100% load. Figure V-2 gives the detailed schematic showing each flow stream. Water removed from the system includes evaporation to the gas and water in the filter cake. Water in the cake includes both free water and chemically combined water (water of crystallization). Water is added to the system in the lime slurry (both as slurry water and chemically combined), cake wash, instrument purge, and pump seals. All water fed to the pump seals enters the system; that which drains out of the seals is collected in sumps and returned to the thickener. The evaporation to the gas is assumed to vary directly with the boiler load and the variations in the inlet gas temperature and inlet gas humidity. The nonlinear relationship between the gas flow rate and the percent load is not considered here. In general, such considerations would tend to increase evaporation of water to the flue gas at low loads over that estimated here, and thereby provide more flexibility in the water balance.

In addition to these water streams, which are directly related to the process operation, there is evaporation of water to the atmosphere from the open vessels (thickener, thickener hold tank, etc.) and addition of water to the system from rainfall. The evaporation to the atmosphere from open vessels is based upon prevailing meteorological conditions in the Louisville area. Mean monthly values of temperature and humidity have been used to determine the range of evaporation. The rainfall corresponds to a range from nil to a daily maximum of 7" per day (34 gpm based on the thickener and thickener hold tank cross-sectional areas).

The noncontact water streams are the booster fan cooling water and the seal water for the vacuum pumps. The noncontact water streams are not included in the process water balance since this water will be segregated from the process streams. Because the noncontact water will contain no pollutants and will have a temperature rise of only 5-10°F, it is expected that the water can be discharged directly in the river.

b. Water Balance Sensitivity

In Figures V-3 and V-4 the total water removed from the system is plotted against water added to the system. The evaporation to the atmosphere from open tanks and rainfall are not included here since they are not controlled as part of the system and are considered perturbations. The noncontact water streams are also not included. The two figures show the water balance for combustion of coal containing 4.7% sulfur and 3.8% sulfur. The expected operating conditions at Louisville are 3.8% sulfur coal, 55% solids in the cake, and a wash ratio of 2.0 (Figure V-4). The capacity of each filter is about 400 lbs of dry cake/min. at the design conditions.

The water balance line (line at 45° to axis) is the locus of points where the water removed from the system is equal to water added to the system. Points falling below and to the right of the water balance line indicate that the system is out of balance, and the points falling above the water balance line show the system is within balance (the balance maintained by adding necessary makeup water to the system). The solid lines representing 100% load, 60% load, and 20% load have points representing 0, 1, 2, or 3

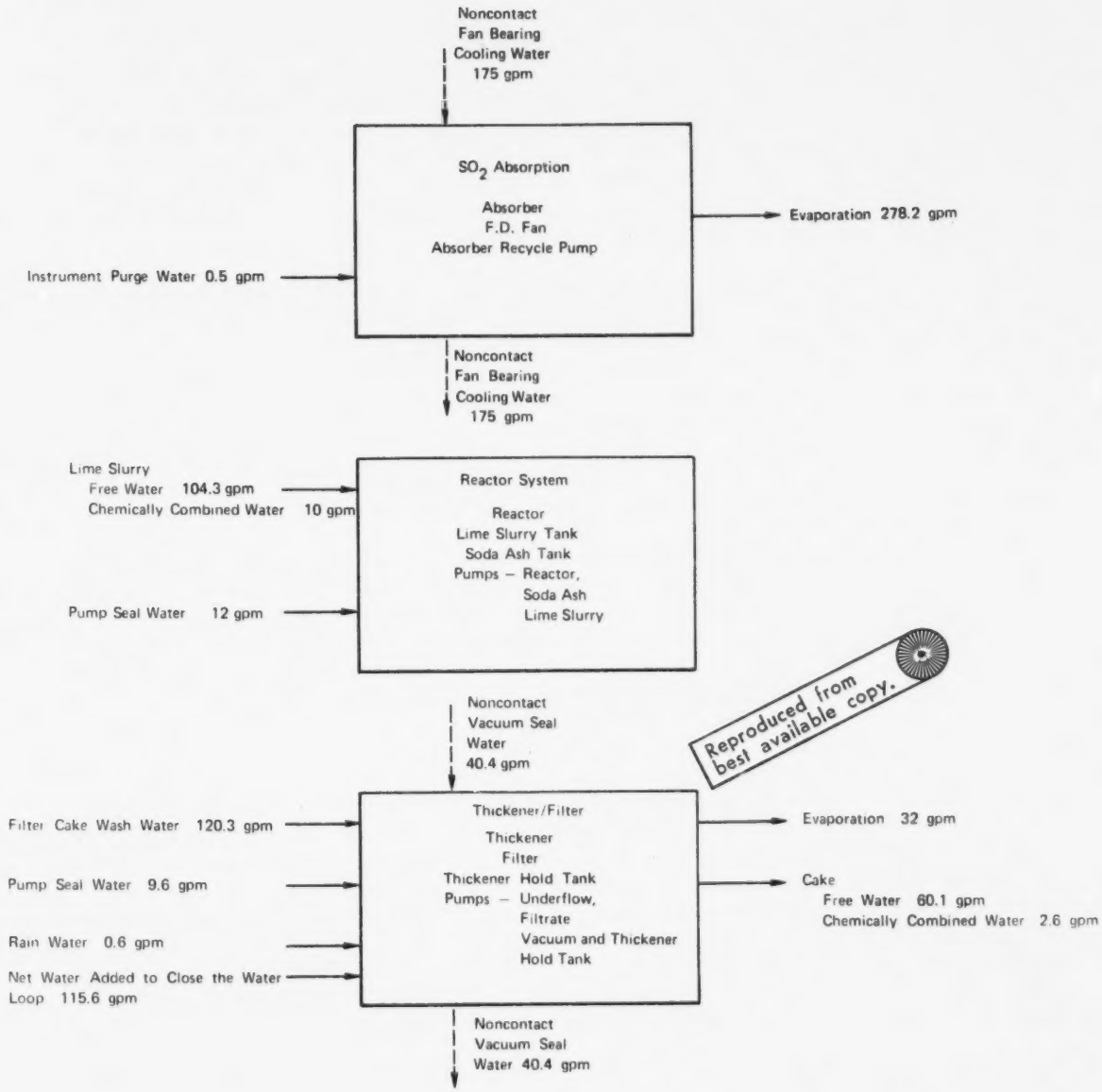

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FIGURE V–1 BLOCK DIAGRAM OF WATER BALANCE AT 100% LOAD

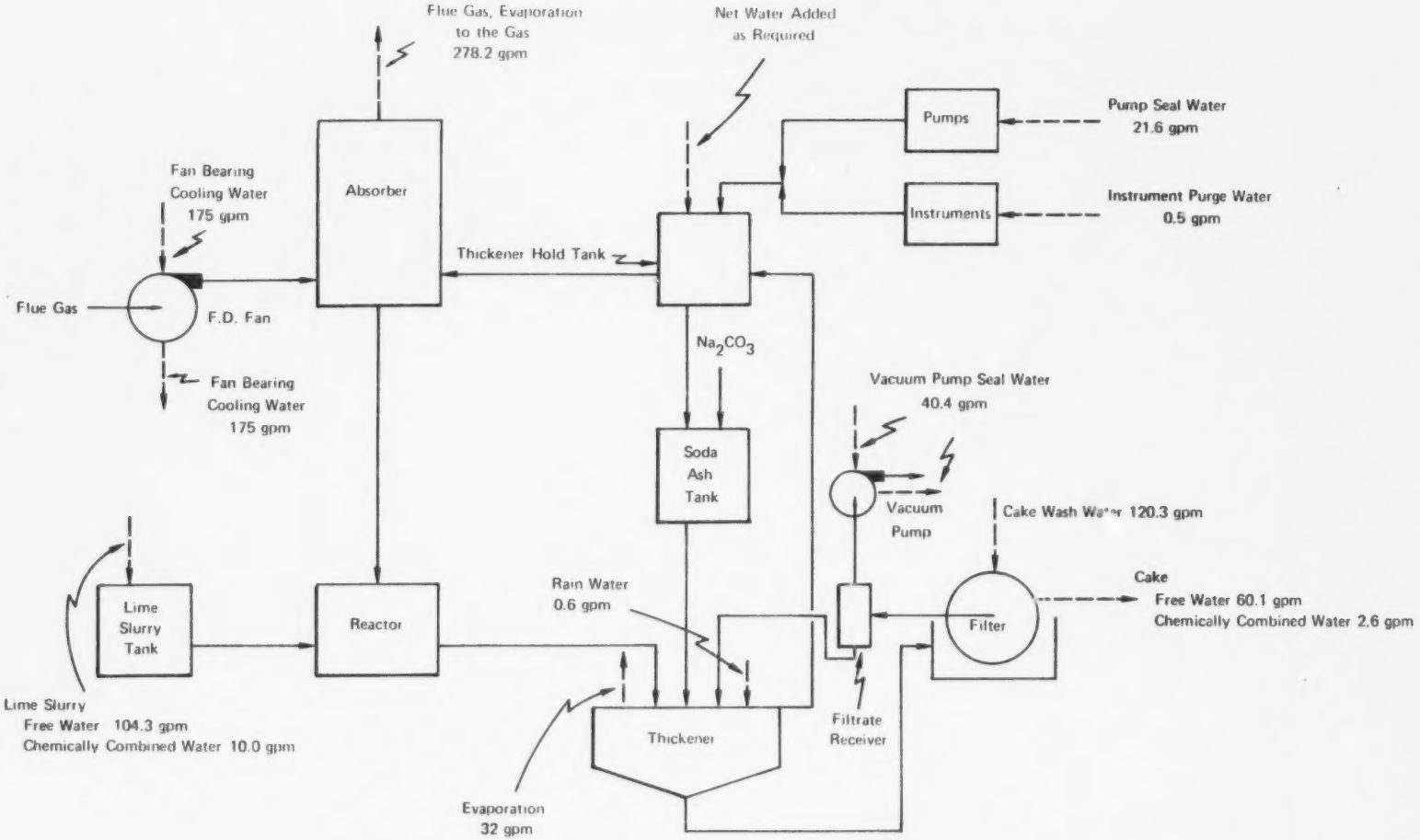
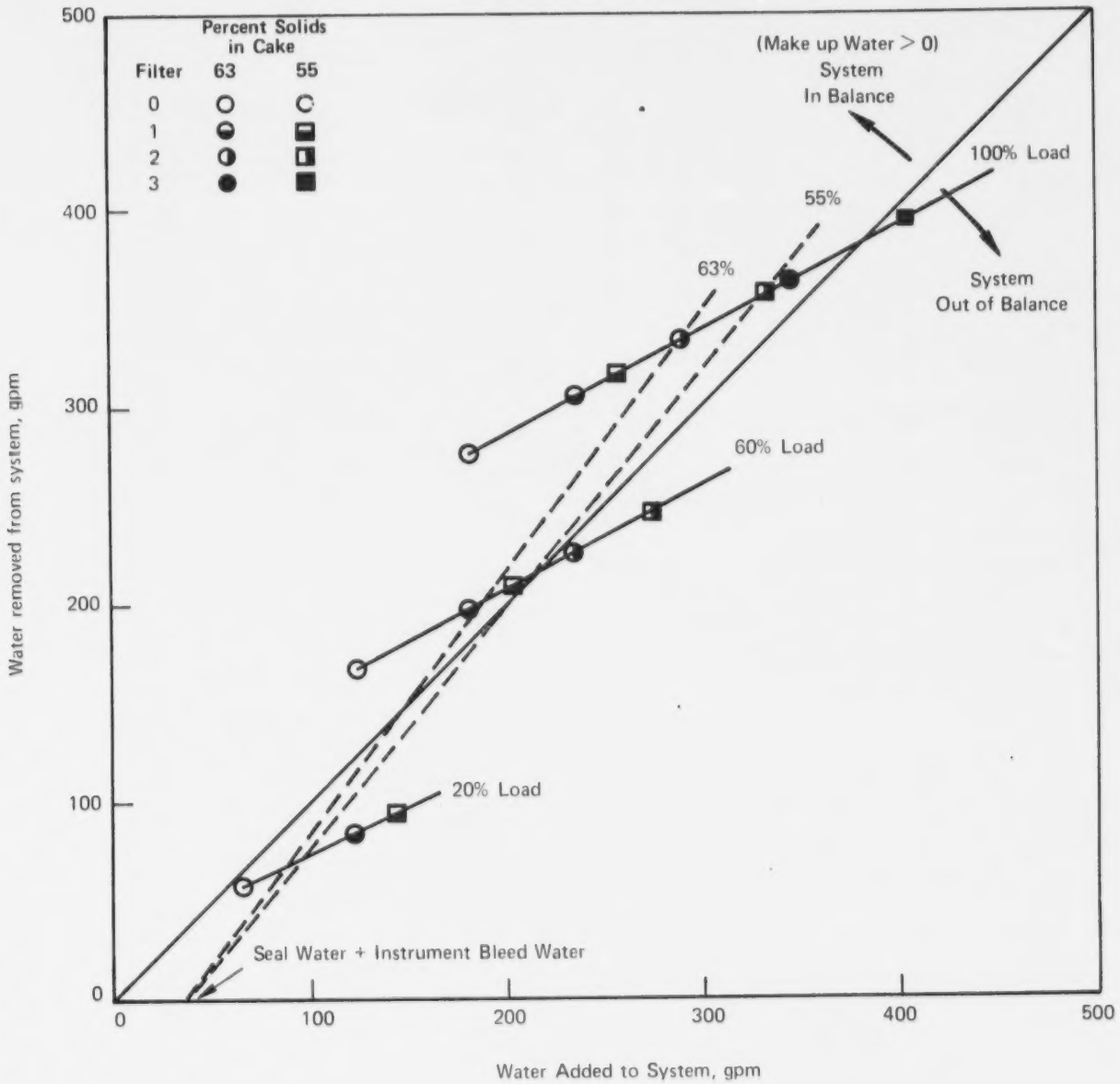


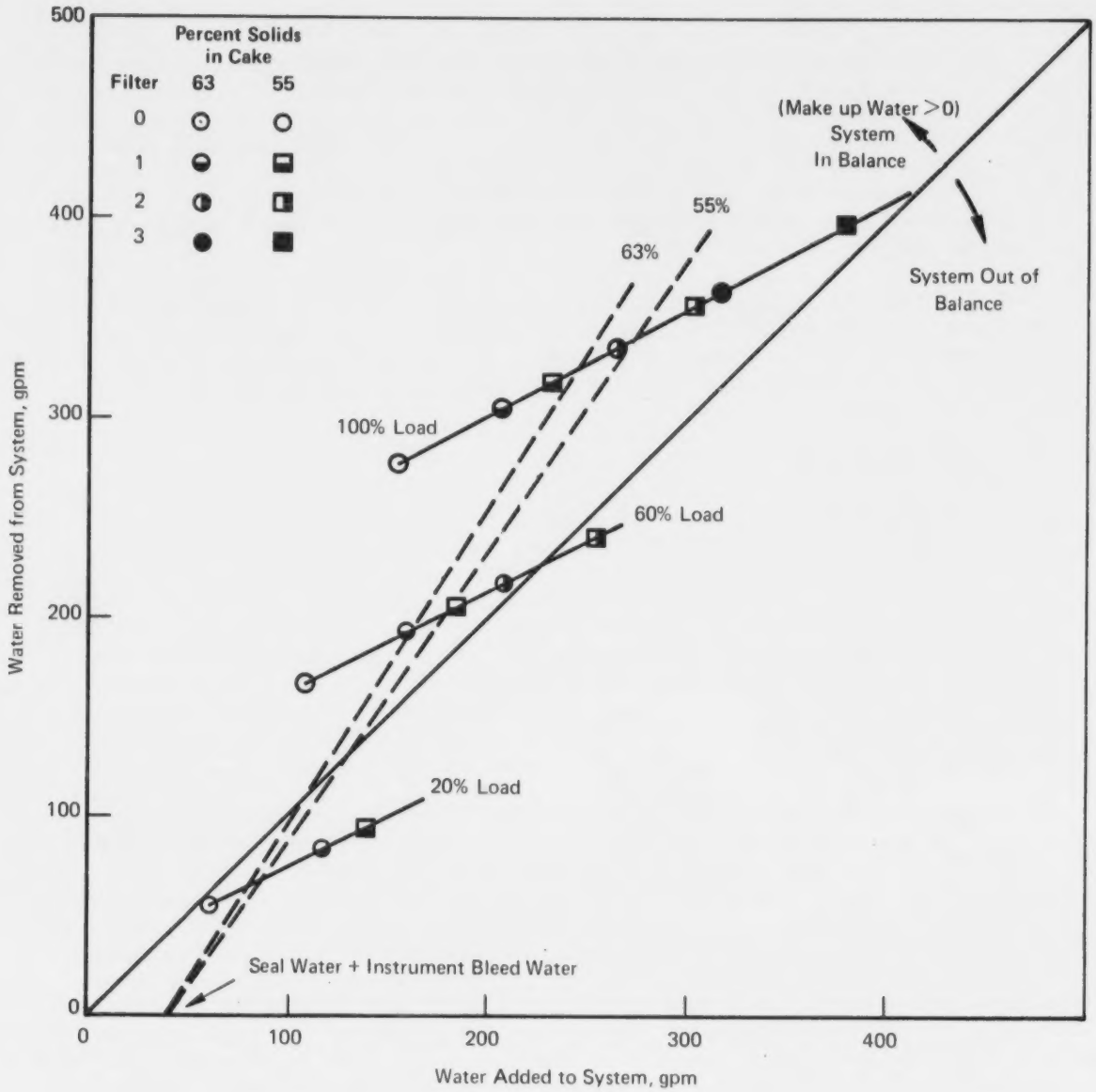
FIGURE V–2 WATER BALANCE AT 100% LOAD (3.8% S COAL, 55% SOLIDS IN CAKE, 2.0 WASH RATIO)
DASHED LINES SHOW WATER INPUT OR WATER OUTPUT FROM THE SYSTEM



Basis: 4.7% S in coal (design condition)
100% seal water added to system (40.5 gpm)
2.0 wash ratio.

Note: Dashed lines show solids generated equal to solids removed for different percent solids in cake.

FIGURE V–3 WATER BALANCE FOR 4.7% SULFUR COAL



Basis: 3.8% S in coal (operating condition)
100% seal water added to system (40.5 gpm)
2.0 wash ratio.

Note: Dashed lines show solids generated equal to solids removed for different percent solids in cake.

FIGURE V–4 WATER BALANCE FOR 3.8% SULFUR COAL

filters in operation. The dashed lines on the plot show solids generated in the system equal to the solids removed from the system for cakes containing different percent solids. Therefore, the dashed lines show steady-state conditions over a long term, and the points (at 20%, 60%, and 100% load in the figures) show the actual operating conditions at a given time. The dashed lines meet this axis at 40.5 gpm (Figures V-3 and V-4), which represents maximum pump seal water (40 gpm) and instrument bleed water (0.5 gpm) added to the system.

The water balance is affected by the number of filters in operation. At low loads (20-35% load) the filters may be shut down to decrease the net water addition to the system. Therefore, solids will be added to the system and the total system volume increase will be the sum of net water and the net solids added to the system.

c. Evaporation and Rainfall

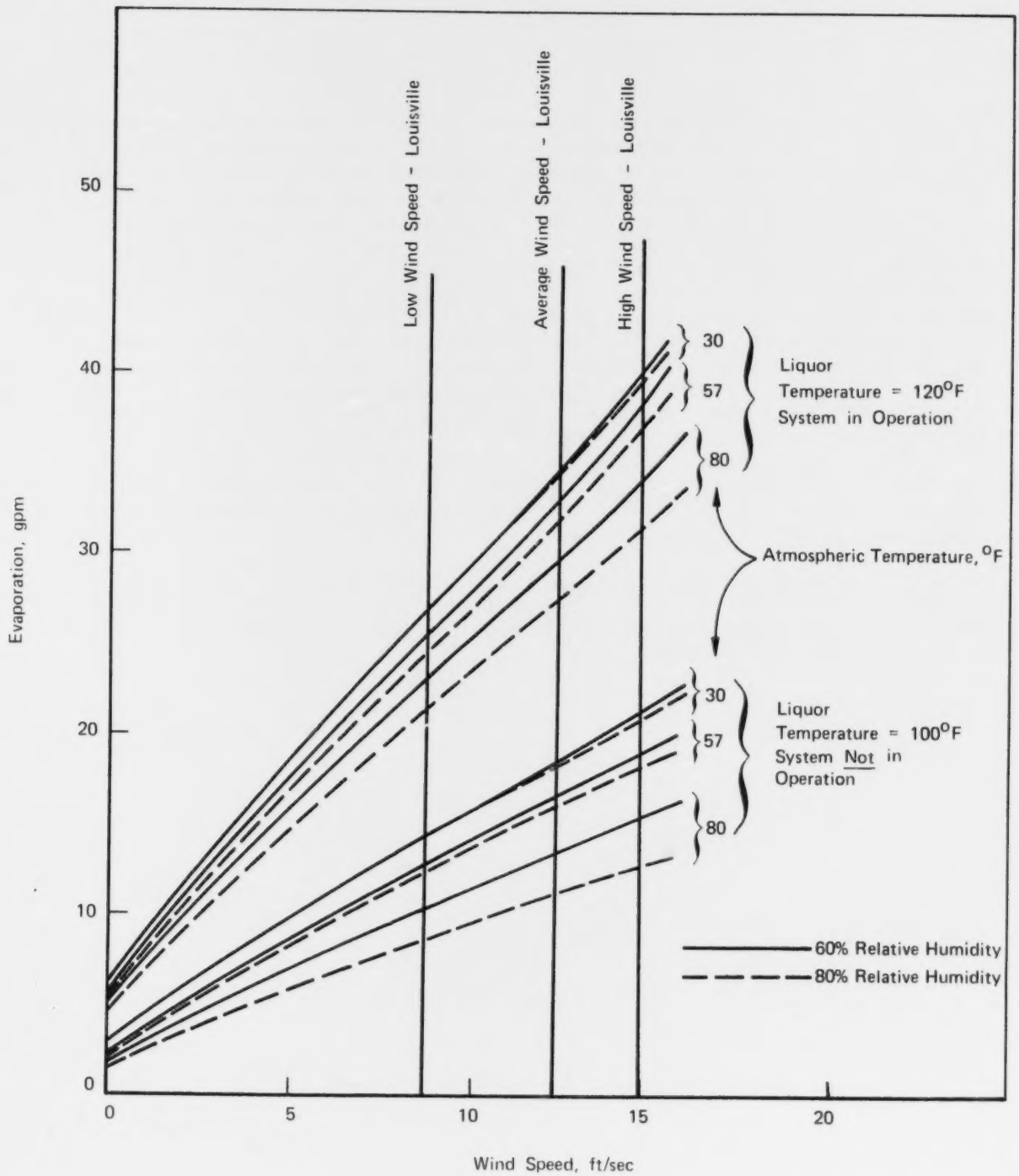
The evaporation of water from the thickener and hold tank is shown in Figure V-5 (total cross-sectional area is equivalent to a cross-sectional area of 120' diameter tank). The evaporation rate is plotted against wind speed. The relationship may be considered to be good to $\pm$25% and is based on the assumption of still liquid surface. In practice, the liquid surfaces are protected by the sides of the vessels, and therefore the average wind speed obtained from meteorological data may not be the same as the wind speed on the liquid surface. Also, the liquid surfaces are not smooth, and therefore the actual surface exposed to the atmosphere is greater than the calculated surface area from the vessel dimensions.

The meteorological conditions shown in Figure V-5 represent the average conditions at Louisville. The mean monthly temperature varies from 30°F to 80°F, and the mean yearly temperature is 57°F. Similarly, the mean monthly wind speed varies from 8.8 ft/sec. to 14.7 ft/sec., and the mean yearly wind speed is 12.4 ft/sec. The humidity varies from 60% to 80%. The liquor temperature in the thickener (at exposed surface) is expected to be 120°F during the operation of the system and lower during shutdown periods. The evaporation rate varies from 21 to 40 gpm (during system operation) with an average evaporation rate of 32 gpm.

The maximum rainfall is expected to be 2"/hour (hourly average), equivalent to a water addition rate of 234 gpm (in the thickener and hold tank). The daily maximum rainfall is 7"/day, equivalent to a water addition rate of 34 gpm. The average rainfall at Louisville is 43"/year, and the average snowfall is 17.5"/year. The total precipitation is equivalent to a water addition rate of 0.6 gpm.

B. ENERGY CONSUMPTION

The energy required for the dual alkali system includes both electrical energy and fuel for reheat. The flue gas supplied to the dual alkali system represents a waste stream, and the difference in the energy content of the outlet gas stream and the inlet gas stream is not considered here.



Note: Mean temperature at Louisville = 57°F.

FIGURE V–5 EVAPORATION IN A 120' DIAMETER THICKENER

The electrical energy requirement for the dual alkali system is presented in Table V-4. The energy requirements in Table V-4 are estimated for the dual alkali system installed on a boiler operating at peak capacity of 300 Mw. The electrical energy is used for fans, pumps, agitators, etc. Equipment consuming little energy, such as lighting, electrical energy for the compressor to supply pressurized air for instruments, and maintenance equipment, are included under miscellaneous items. The total electrical energy requirement is estimated to be 3.1 megawatts, or approximately 1.0% of the peak power generation for Unit No. 6. About 60% of the total electrical energy is required for the booster fans, 10% for reheater fans, and 30% for the rest of the system. The energy requirement for the fan varies in direct proportion to the design gas flow rate and therefore varies directly with the boiler capacity. (The pressure drop through the system is assumed to be constant for this purpose.) The 30% of the energy used in the dual alkali system varies roughly with the feed forward rate in the system. The feed forward rate is in direct proportion to the sulfur dioxide removed from the system. (The liquor concentration is assumed to be fixed for this purpose.)

No. 2 fuel oil will be used to reheat the saturated gas. The wet gas will be reheated from 126°F to 176°F. The estimated fuel requirement is 343 gal/hour when the boiler is generating 300 Mw. The fuel oil energy requirement is equal to 48×10^6 Btu/hour.

TABLE V-4

ELECTRICAL ENERGY REQUIREMENTS FOR THE DUAL ALKAKI SYSTEM
AT THE LOUISVILLE GAS AND ELECTRIC COMPANY

Equipment	Number at Full Load		BHP	IHP	Total BHP At Full Load
	Operating	Spare			
Booster Fan	2	0	1,246	1,250	2,492
Pumps:					
Absorber Recycle	2	2	215	250	430
Thickener Hold Tank	1	1	157	200	157
Thickener Underflow	1	1	33	40	33
Soda Ash	1	1	6	10	6
Vacuum	1.5	1.5	100	100	150
Reactor	2	0	59	75	118
Filtrate Sump	1	1	20	25	20
Lime Slurry	1	1	30	30	30
Underflow Sump	0	2	8.5	15	0
Soda Ash Sump	0	2	5	7.5	0
Agitators:					
Lime Slurry	1	0	20	20	20
Primary Reactor	2	0	7.5	7.5	15
Secondary Reactor	2	0	25	25	50
Soda Ash	1	0	1.5	1.5	1.5
Vacuum Filter	1.5	1.5	1.5	1.5	2
Thickener:					
Mechanism	1	0	5	5	5
Rake	0	1	3	3	0
Filter:					
Drum Drive	1.5	1.5	5	5	7.5
Blowback Fan	1.5	1.5	2	2	3
Soda Ash Weigh Feeder					
Bin Vibrator	1	0	1.5	1.5	1.5
Weigh Feeder	1	0	1	1	1
Dust Collector	0	1	0.5	0.5	0
Bin Slide Gate	0	1	0.5	0.5	0
Reheater					
Primary Fan	2	0	100	100	200
Secondary Fan	2	0	40	40	80
Compressor	2	0	7.5	7.5	15
Miscellaneous: Instrumentation, Space Heaters, Lighting, Bearing Heaters, etc.			200	200	200
				TOTAL hp	4,037.5
				TOTAL kw	3,102
				% of 300 Mw	1.0%

VI. PROCESS EQUIPMENT AND INSTRUMENTATION

A. EQUIPMENT

1. General Arrangement

The plot plan for the dual alkali system is shown in CEA drawing[a] 040044-2-0. The general arrangement for individual equipment is shown in the following CEA drawings.[a] These drawings are preliminary and if required will be updated in the Phase II report to reflect later modifications.

Absorber Lower Plan (CEA drawing 040044-2-1)

The ducting connected to the induced draft fan of the boiler system, booster fan for the dual alkali system, absorber bypass ducting, chimney base, recycle pumps and the pump building for the absorbers are shown in the drawing. The dampers indicated by DJ, platforms and ladders in the lower part of the absorber are also shown here. The drawing shows only absorber A101 section. Absorber A102 section is the mirror image and is shown in CEA drawing 040044-2-3. Absorber sections A-A and B-B indicated for absorber A101 are shown in CEA drawings 040044-2-8 and 040044-2-9, respectively.

Absorber A101 Upper Plan (CEA drawing 040044-2-2)

The drawing shows the upper part of the absorber, outlet ducting, reheater, connection of ducting to breeching and the chimney. The two manholes in the upper part of the absorber, platform, and ladders are also shown here. The drawing shows absorber A101 section only. Absorber A102 section is the mirror image of this drawing and is shown in CEA drawing 040044-2-4.

Reactor, Silo, Vacuum Filter Areas Plan (CEA drawing 040044-2-6)

The plan is shown at two elevations--at grade and at elevation 475 ft. The primary and secondary reactors, reactor pumps, filters, filtrate receivers, vacuum pumps for filter, lime slurry tanks, lime slurry pumps, soda ash silo, weigh feeder, soda ash solution tank and soda ash solution pumps are all shown in this drawing. There are two reactor systems, three filter systems, two lime slurry pumps and two soda ash solution pumps. Also, part of the thickener and thickener hold tank are shown in this drawing. Sections C-C, D-D and E-E indicated in this drawing are shown in General Arrangement Sections-Reactor and Silo Areas, Vacuum Filter and Pump Building (CEA drawing 040044-2-11).

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[a] A set of large size CEA drawings may be obtained from Environmental Protection Agency, Research Triangle Park, North Carolina. Because of too many details, these drawings cannot be reduced; and, therefore, are not included here.

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Thickener, Pipe Tunnel and Filter Building (CEA Drawing 040044-2-10)

Sections A-A and B-B from the Thickener and Hold Tank Plan (CEA drawing 040044-2-5) are shown here. The drawing shows the center well for the thickener, the superstructure, pipe rack to the feed well, the thickener underflow pumps, the pipe tunnel and accessway underneath the thickener.

2. Equipment Details

Table VI-1 contains a list of equipment in the system. Materials of construction for the system are summarized in Table VI-2.

The system is designed with appropriate corrosion resistance where required using stainless steel (316 or 317) or linings (polyester or rubber). The expected chloride levels in the process liquor range from 10,000 ppm to 15,000 ppm; but, levels can vary from as low as a few thousand ppm to almost 20,000 ppm depending upon the chloride content of the coal and the degree of cake washing. Liquor pH's range from about 5.0 in the absorber loop to greater than 12.0 in the reactor and dewatering systems.

With the exception of the primary reactors, all tanks and vessel linings in contact with process liquor are lined with flake reinforced polyester. The primary reactor is constructed of 316L ss; the filtration equipment is both 316L ss and fiberglass; and the absorber trays are 317L ss.

All pumps and agitators in contact with process liquor are rubber lined. Process liquor piping is FRP. Hot flue gas ducting is carbon steel; and the booster fan housing and blades are A441 steel. Saturated flue gas ducting is polyester lined between the absorbers and reheater section; and is 317L ss between the reheater and stack.

Tables VI-3 through VI-14 give details regarding the design basis, service, dimensions, and materials of construction for each of the equipment items listed in Table VI-1.

In accordance with standard design practice at LG&E, all the tanks in Table VI-10, except the thickener hold tank, are closed top.

B. PIPING AND INSTRUMENTATION

1. General Description

The preliminary list of instrumentation for the dual alkali system is shown in Table VI-15, and the piping and instrument diagrams are given in CEA drawings[a] 040044-1-3, 4 and 6. The system is designed to minimize operator interface. All the remote controls are located in a centralized control room from which the system can be started up, operated, and shut down. The control room is furnished with appropriate controllers, indicators, recorders, switches, and other necessary instrumentation for the safe and convenient operation of the system. Included on the control panel are:

[a] A set of large size CEA drawings may be obtained from Environmental Protection Agency, Research Triangle Park, North Carolina. Because of too many details, these drawings cannot be reduced; and therefore are not included here.

TABLE VI-1

EQUIPMENT LIST

Equipment	Description	No. Required
1. Agitators (w/motors)	Reactant Feed Tank	1
	Primary Reactor Tank	2
	Secondary Reactor Tank	2
	Soda Ash Solution Tank	1
	Vacuum Filter	3
2. Dampers	Booster Fan Inlet	2
	By-Pass	1
	Outlet	2
	Reheater Fan	2
3. Ductwork	Take-Off Connecting Duct	2
	Booster Fan Inlet Duct	2
	Booster Fan Outlet Duct	2
	Absorber Outlet Duct	2
	Duct/By-Pass Transition	1
4. Expansion Joints	Tie-in	2
	Booster Fan Inlet	2
	Booster Fan Pant Leg	4
	Booster Fan Outlet	2
	Booster Fan Outlet Duct	2
	Absorber Outlet PC#1	2
	Absorber Outlet PC#2	2
	Breeching	1
	By-Pass	2

TABLE VI-1
(Cont.)

Equipment	Description	No. Required
5. Fans	Booster Fan, Drive	2
	and Motor	2
	Reheater Fan & Motor	2
	Filter Blow Back Fan & Motor	3
6. Heaters	Reheater	2
7. Pumps	Reactant Feed Pump & Motor	2
	Reactor Transfer Pump & Motor	2
	Thickener Underflow Pump & Motor	2
	Soda Ash Solution Pump & Motor	2
	Hold Transfer Pump & Motor	2
	Service Pump & Motor	1
	Absorber Recycle Pump & Motor	4
	Vacuum Pumps & Motor	3
	Filtrate Sump Pumps & Motor	2
	Soda Ash Sump Pump	2
	Underflow Sump Pump & Motor	2
8. Tanks	Reactant Feed Tank	1
	Primary Reactor Tank	2
	Secondary Reactor Tank	2
	Soda Ash Silo & Accessories	1
	(1) Dust Collector	1
	(1) Bin Bottom with Motor	1
	(1) Vibrator with Motor	1
	Soda Ash Delumper & Motor	1
	Weigh Feed & Motor	1
	Thickener Hold Tank	1
	Soda Ash Solution Tank	1
	Filtrate Tanks	3

TABLE VI-1
(Cont.)

	Equipment	Description	No. Required
9.	Tanks (Cont.)	Thickener	1
10.	Thickener Mechanism	Thickener Mechanism & Motor	1
		Thickener Lift Rake Motor	1
11.	Vacuum Filter	Vacuum Filter System & Drive	3
12.	Vessels	Absorber	2

TABLE VI-2

MATERIALS OF CONSTRUCTION

1. Agitators

   (a) Lime slurry tank--carbon steel.

   (b) Primary reactors--shaft: carbon steel, rubber lined; hub and blades: 317L s.s.

   (c) Secondary reactors--shaft: carbon steel, rubber lined; hub and blades: 317L s.s.

   (d) Soda ash tank--carbon steel, rubber lined.

2. Dampers

   (a) Booster fan inlet --A36 steel

   (b) Bypass and absorber outlet--316L s.s.

3. Duct Work

   (a) Duct work carrying hot flue gas to the scrubber inlet is carbon steel.

   (b) Duct work carrying saturated flue gas from the scrubber to reheater is carbon steel, coated internally with a flake reinforced polyester lining.

   (c) Duct work carrying reheated flue gas from the reheater to stack is 317L s.s.

   (d) The bypass/transition duct is carbon steel.

4. Expansion Joints

   (a) Expansion joints on inlet side of absorbers are of viton.

   (b) Expansion joints on outlet side of absorbers are of chlorobutyl for wet gas and of viton for dry gas (after reheat).

5. Fans

   (a) Housing--A441.

   (b) Blades--A441, with wear plates constructed from A441 material.

TABLE VI-2
(Cont.)

6. Pumps

   (a) Housing--rubber lined

   (b) Impeller--rubber lined

7. Tanks

   (a) Thickener overflow tank--carbon steel, flake reinforced polyester lining.

   (b) Primary reactors--316L s.s.

   (c) Secondary reactors--carbon steel, flake reinforced polyester lining, rubber pad on bottom.

   (d) Lime slurry storage tank--carbon steel.

   (e) Soda ash mixing tank--carbon steel.

8. Soda Ash Silo

   (a) Carbon steel

9. Weigh Feeders

   (a) Frame--mild steel

   (b) Internals--304 s.s.

10. Thickener

   (a) Thickener shell--concrete with carbon steel interior, flake reinforced lining.

   (b) Rake, shaft and centerwell--carbon steel, rubber lined.

11. Vacuum Filter

   (a) Filter drum--316 s.s.

   (b) Agitator--316 s.s.

   (c) Filtrate receiver--FRP

TABLE VI-2
(Cont.)

12. Absorber

(a) Absorber shell--carbon steel, coated internally with flake reinforced polyester lining.

(b) Absorber trays--317 s.s.

(c) Demister--polypropylene.

13. Piping

(a) All process piping is FRP.

(b) Piping for make-up water and service water and all other piping not subject to corrosion is carbon steel.

TABLE VI-3

AGITATORS

	Lime Slurry Storage Tank	Reactor 1	Reactor 2	Soda Ash Tank
Number Required:	1	2	2	1
Impeller Type	turbine	turbine	turbine	propeller
Impeller dia.	87"	62"	92"	12"
RPM	30	45	37	350
Shaft dia.	4.5"	3"	5"	1.5"
H.P.	20	7.5	25	1.5
Material of Construction:				
Shaft	C.S.	R.C.C.S.	R.C.C.S.	R.C.C.S.
Blades	C.S.	317L S.S.	317L S.S.	R.C.C.S

Data are given per agitator
C.S. - Carbon steel
R.C.C.S - Rubber covered carbon steel
S.S. - Stainless steel

TABLE VI-4

DAMPERS

	Booster Fan Inlet	Bypass	Absorber Outlet
Number Required:	2	1	2
Design flow rate - ACFM - °F	 533,000 350	 1,065,000 350	 487,000 200
Size	135-1/4" x 138-1/2"	162" x 240"	156" dia.
Position of duct	Horizontal	Horizontal	Horizontal
Type	Guillotine	Multi-louver	Guillotine
Entry	Bottom	--	Bottom
Normal position	Open	Closed	Open
Material of Construction	A36 steel	316L s.s.	316L s.s.
Max. gas leakage, % of design flow rate	1	1	1
Paint external members	Zinc chromate	Zinc chromate	Zinc chromate

TABLE VI-5

DUCTWORK

Equipment	No. Required	Dimensions
Take off connecting duct	2	Inlet 11'8" x 11'6" Outlet to booster fan inlet duct 11'3" x 11'6" Outlet to bypass transition 11'1" x 12'1" Overall dimensions 11'8" x 11'6" x 16'10"
Booster Fan Inlet Duct	2	Inlet 11'3" x 11'6" Outlets16'8" x 3'2" (two) Overall dimensions 11'6" x 28' x 16'6"
Booster Fan Outlet Duct	2	Inlet duct 10'7" d Outlet duct 13'9" d Overall length 12'
Absorber Outlet Duct	2	Inlet 13'd Outlet 13'd Overall length 80'
Duct Bypass/Transition	1	Inlets from take off connecting duct 11'1" x 12'1" (two) Inlets from absorber outlet duct 13'd (two) Outlet 28' x 13'6" Overall dimensions 30'3" x 24'6" x 19'10"

See Table 7.3 for materials of construction.

TABLE VI-6

EXPANSION JOINTS

Service	No. Required	Size	Material
Tie-In	2	11'10" x 11'8-1/2"	Viton
Booster Fan Inlet	2	11'10" x 11'8-1/2"	Viton
Booster Fan Pant Leg	4	3'4-3/8" x 16'10-1/2"	Viton
Booster Fan Outlet	2	10'9-7/8" x 8'10-1/8"	Viton
Absorber Inlet	2	13'11-5/8" dia.	Viton
Absorber Outlet 1	2	13'2" dia.	Chlorobutyl
Absorber Outlet 2	2	13'2" dia.	Viton
Breeching	1	13'8" x 28'2"	Viton
Bypass	2	11'10" x 11'8-1/2"	Viton
Reheater Duct	2	Reheater vendor to specify size and material	

All the expansion joints except absorber outlet 1 and reheater duct expansion joints are designed for 400°F with excursions to 600°F for 5 minutes at 4 times a year. All the expansion joints are 9" wide.

TABLE VI-7

BOOSTER FANS

Number Required	2
Flue gas Volume	533,000 acfm
Flue gas temperature	300°F
Inlet static pressure	+2 inch WG
Outlet pressure at design flow rate	10.5 inch WG
Gas density	0.0526 lb/cu ft
Inlet dust loading	0.0537 gr/cu ft
Maximum vibration amplitude	2.2 mils at 720 rpm
Type	Centrifugal forced draft
Fan blade design	Backward inclined airfoil with wear plates
Materials of construction	Carbon steel
Drive	Fluid drive
Motor HP	1,250
Volts	4,000
Motor rpm	720

TABLE VI-8

REHEATERS

Operating Conditions	
Wet flue gas flow rate	436,500 acfm
Temperature	126°F
Pressure	+2 inch WG
H_2O Vapor	2,475 lbs/min
SO_2	11.25 lbs/min
	200 ppm dry basis
Particulates	2.48 lbs/min
Temperature after reheat	176°F
Heater Requirements	
For Flue gas	25,632,000 Btu/hr
Radiation loss	1,282,000 Btu/hr
Heater outlet temperature	800°F max.
Turndown	to 20% of the capacity
Heater outlet pressure	+7 inch WG
Fuel Data	
Fuel type	Number 2 fuel oil
Oil flow rate	171 gal/hr
Air Inlet Temperature	
Winter	0°F
Summer	100°F
No. of Reheaters Required	2

TABLE VI-9

PUMPS

	Absorber Recycle Pump	Reactor Pump	Thickener Hold Tank Pump	Thickener Underflow Pump	Lime Slurry Pump	Soda Ash Pump
Number Required						
Operating	2	2	1	1	1	1
Spare	2	-	1	1	1	1
Capacity						
gpm	4,600	1,965	4,185	665	340	140
head, ft	130	85	105	115	115	75
Speed	Variable	Variable	Variable	Variable	Variable	Variable
Material of Construction	RLCI	RLCI	RLCI	RLCI	NI-Hard	RLCI
Packing	Yes	Yes	Yes	Yes	Yes	Yes
Drive	V belt	V belt	V belt	V belt	V belt	V belt
Motor Mounting	Overhead	Overhead	Overhead	Overhead	Overhead	Overhead
Voltage, volts	4000	460	4000	460	460	460
Drip proof	Yes	Yes	Yes	Yes	Yes	Yes
BHP/IHP	215/250	62/75	157/200	33/40	30/30	6/10
Service Factor	1.15	1.15	1.15	1.15	1.15	1.15
Overall Size[a]	12x10x25	10x8x21	12x10x25	5x5x14	3x1.5x16	1x1.5x6

RLCI - Rubber lined cast iron
All pumps are centrifugal pumps.
a - All dimensions are in feet.

TABLE VI-10

TANKS

Process Data	Primary Reaction Tank	Secondary Reaction Tank	Thickener Hold Tank	Lime Slurry Storage Tank	Soda Ash Tank
Liquor specific gravity	1.1	1.1	1.1	1.2	1.1
pH range	5-11	11-12.5	12	12	12
Chlorides, ppm	12,000	12,000	12,000	---	12,000
Operating pressure, in wg	Liquid head	Liquid head	Liquid head	Liquid head	Liquid head
Design pressure, in wg	Liquid head	Liquid head	Liquid head	Liquid head	Liquid head
Operating temperature, °F	126	126	110	70	110
Design temperature, °F	---	---	---	100	---
Specified data					
Minimum thickness, inches	---	1/4	1/4	1/4	3/16
Seismic zone	1	1	1	1	1
Code	API650	API650	API650	API650	API650
Tank shape	Cylindrical	Cylindrical	Cylindrical	Cylindrical	Cylindrical
Dimensions, dia x height	11' x 14'	20' x 33'	36'x 23'	24'6" x 24'6"	6' x 8'
Baffles	4	4	None	4	4
Agitator	Yes	Yes	No	Yes	Yes
Materials					
Shell and head	316L S.S.	A283 (C.S.)	A283 (C.S.)	A283 (C.S.)	A283 (C.S.)
Internal structure	316L S.S.	C.S.	C.S.	C.S.	C.S.
Nozzles necks/flanges	316L S.S.	C.S. & 316L S.S.	C.S. & 316L S.S.	C.S.	C.S. & 316L S.S.
Lining	None	Glass reinforced polyester + 3/8" thick rubber pad on bottom	Glass reinforced polyester	---	Glass reinforced polyester
Paint	Zinc chromate primer	Zinc chromate primer	Zinc chromate primer	Zinc chromate primer	Zinc chromate primer
Gaskets	Neoprene	Neoprene	Neoprene	Neoprene	Neoprene
Erection Weight, lbs	75,000	36,000	39,000	39,000	2,500
Operating Weight, lbs	97,500	71,500	1,183,000	785,000	17,500

C.S. = carbon steel; S.S =stainless steel; all tanks are closed at top except thickener hold tank.

TABLE VI-11

SODA ASH SILO

Process Data	
Specific gravity	65 lbs/ft^3
Soda ash type	dense
Specified Data	
Minimum thickness	1/4"
Wind Load at 30'	30 PSF
Seismic Zone	1
Code	API650
Tank shape	Cylindrical plus conical bottom
Size	12' diameter x 34' 6" height cylinder 6' bottom diameter x 5'6" height cone
Thickness	3/8" thick bottom of cylinder
	5/16" thick middle of cylinder
	1/4" thick top of cylinder
	1/4" thick cone
Attachments	Cone vibrator, baghouse at top, and piping at top for pneumatic feed system
Materials	
Shell and head	A36 (C.S.)
Internals	C.S.
Nozzles Necks/flanges	C.S.
Paint	Zinc chromate primer
Gaskets	Neoprene
Erection Weight	22,000 lbs.
Operating Weight	300,000 lbs.

TABLE VI-12

THICKENER

Design stream conditions	
Reactor Bleed	3,570 gpm
Filter overflow	148 gpm
Filtrate	500 gpm
Soda ash solution	50 gpm
Thickener recycle	155 gpm
Thickener underflow	605 gpm
Operating temperature	110° F
Seismic zone	1
Thickener type	Flat bottom
Diameter	125'
Height	23'
Feed well	
Diameter	12'
Height	11'
No. of arms on rake	2
Cone scrapper	Yes
Overflow weir plate	Notched
Access walk way	On one side of the superstructure
Rake drive motor	5 HP
Rake lifting device motor	3 HP
Materials	
Rake	Rubber covered carbon steel
Feed well	Rubber covered carbon steel
Shell	Concrete lined with carbon steel
Bottom	Concrete
Weir Plate	FRP
Shell and Bottom	Lined with glass reinforced polyester
Paint	Zinc chromate primer

TABLE VI-13

VACUUM FILTER

Process Data	
Net slurry feed	Design 3,147 lbs/min. 302 gpm
	Maximum 4,533 lbs/min 448 gpm
Specific gravity	1.25
Temperature	140°F maximum
pH	11-12.5
Wash water	2,108 lbs/min design
Filter Requirements	
% Solids in cake	55% minimum 63% average
Total flow	1,246 lbs/min design 1,584 lbs/min maximum
Filtrate	
Flow rate	299-750 gpm
pH	10.5-12.3
Specific gravity	1.064
Cake wash rate	91 gpm normal 300 gpm maximum
No. of filters required	3 including spare
Size	8' diameter 10'face
Cake discharge mechanism	Blower assisted scraper blade
Liquid level control in filter tub	Adjustable overflow weir
Cake wash assembly	Drip wash nozzles with drag net
Filter drum speed	0.3-3.0 rpm
Agitator type	Counter weighted rocker arm
Filtrate Receiver	54" diameter x 54" height
Motor	
Vacuum pump	100 HP
Drum drive	5 HP
Filter blow back pump	2 HP
Agitator drive	1.5 HP
Materials	
Filter drum	317 ELC stainless steel
Filter agitator	317 ELC stainless steel
Filter scraper	317 stainless steel with rubber tip
Filtrate receiver	FRP
Drainage grid	Polypropylene
Filter media	Polypropylene

TABLE VI-14

ABSORBER

Process Data	
Specific gravity	1.2
pH range	5-12
Chlorides, ppm	12,000
Operating Pressure, inch WG	+11.5
Design Pressure, inch WG	+12.5
Operating Temperature, °F	125
Design Temperature, °F	350
Upset Conditions	
Temperature, °F	600
Time; minutes	5
Specified Data	
Corrosion allowance	none
Wind Load at 30', PSF	30
Seismic zone	1
Code	API 650
Tank Shape	Cylindrical shell with conical head
Size	32' diameter x 45' height_shell 10'6" height x 13' top diameter_cone
Thickness, inch	3/8
Internals	Sprays + 2 trays + chevron demister
Materials	
Shell and head	A283 (carbon steel)
Trays/supports	317L S.S.
Spray nozzles	316 S.S.
Internal piping/supports	FRP
Mist eliminator/supports	Polypropylene
Nozzle necks/flanges	C.S. and S.S.
Internal fasteners	C.S.
Gaskets	Neoprene
External paint	Zinc chromate
Lining	Glass reinforced polyester
Erection	
Erection weight, lbs.	108,000
Operating weight, lbs.	475,000

TABLE VI-15

PRELIMINARY LIST OF INSTRUMENTS FOR THE DUAL ALKALI SO_2 CONTROL SYSTEM

Type of Instrument	Number
Ammeters for motors	24
Density meter	1
Flow control valves	50
Flow indicators for process solutions	18
Flow indicators for water	21
Flow indicators for purge air	34
Hand switches for manual operation	19
Level indicators	14
Manual indicating controller for damper	1
pH indicators	12
Pressure indicators	22
Soda ash weigh feed recorder	1
Solenoid valves for seal water	18
SO_2 monitors	3
Temperature indicators	14

Gas Side

- SO_2 indicators and recorders monitoring SO_2 concentrations in the inlet flue gas to the absorbers and the outlet flue gas concentration before reheat.

- Temperature indicators for the flue gas at the inlet to the booster fan, at the inlet of the absorber, absorber outlet, and reheated flue gases.

- Pressure indicators on the gas side include pressure of the gas entering the booster fan, pressure at the downstream of the fan, pressure at the absorber outlet, and pressure after the gases are reheated. Pressure taps are provided in the absorber to measure pressure at intermediate points.

- Instruments and controllers necessary for controlling the operation of the fan and the dampers so that the flue gas desulfurization system will not affect the operation of the boiler. The fan and the dampers have been interlocked for emergency shutdown such as if the flue gas temperature in the absorber rises above a specified limit.

Liquid Side

- Status indicators for motors on all pumps, agitators, fans and blowers, filter drum and thickener rake drives, and solenoids operating control valves. Green and red lights are used to indicate the status of motors solenoid valves, etc. A green light is used to indicate the closed, or off position; and a red light is used to indicate the open, or on position.

- pH indicators for indicating pH in the reactor system bleed, absorber bleed and top tray feed.

- Flow indicators and totalizers to measure feed chemicals, makeup water and internal flows.

- Liquid level indicators for the absorber recycle tanks, secondary reactors and thickener hold tank.

2. Special Considerations

Two elements and two transmitters have been included at all critical control points in the system to provide spare instrumentation (on gas side to measure pressures and temperatures and on liquid side to measure levels and pH's.) A selector switch in the control room is used to select the element and the corresponding transmitter.

All the process lines are equipped with sufficient drain and flush valves. The flush valve locations have been carefully selected where slurries are present.

All control valves with the exception of the lime slurry feed can be bypassed so that the system can be operated in case of control valve failure. Bypasses consist of three block valves, one on the bypass line and two around the control valve. A control valve bypass is not provided on the lime slurry line, since a bypass system here would be less useful due to plugging problems. However, the reactor system can be operated at 100% of the capacity with one reactor module for short durations while lime slurry control valves are repaired.

All the pumps are individually connected with seal water supply lines. Flow of seal water to each pump is adjusted locally using a hand control valve and rotameter. Each seal water supply line is also equipped with a solenoid valve interconnected with the pump start/stop switch to automatically ensure a supply of seal water. When the pump is turned on, the solenoid valve opens and the seal water is supplied to the pump. When the pump is turned off, the solenoid valve closes and shuts off the seal water. The solenoid valves are normally closed to prevent water leakage into the system during pump downtime and power failures.

All the pumps are fitted with pressure gauges to indicate the pressures developed by pumps. Similarly, the suction side of all vacuum pumps have vacuum gauges to indicate the vacuum developed.

All pumps have check/block valves which open automatically when the pump is turned on, and close when the pump is turned off. The digital logic to control timings for the valves and pumps is not finalized. Pumps are also equipped with hand switches so they can be operated in the field or from remote location in the control room.

All the motors for the pumps, agitators, etc., have ammeters indicating the current drawn by the motors. The current is indicated locally, but if the current exceeds the preset value, an alarm in the control room is activated.

3. Operating Philosophy

The liquid levels in the various tanks are controlled either by bleed or feed stream flows to these tanks, or by overflow weir position. The control of liquid levels in various tanks is shown in Table VI-16. The primary reactors, thickener, and filter tubs are operated on overflow. The absorber recycle tanks, secondary reactors, and soda ash tank are operated on liquid level control with level maintained by the bleed streams from these tanks. The liquid level in the lime slurry feed tank is maintained by the lime slurry feed rate to the reactant feed tank, and the liquid level in the thickener hold tank is maintained by the makeup water rate to the system.

The feed forward flow rate from the thickener hold tank to absorber is controlled for each absorber individually, to achieve the desired SO_2 removal in the system. The absorber bleed pH provides an indication of the SO_2 concentration in the effluent gas stream and is used to adjust

TABLE VI-16

CONTROL OF STREAM FLOW RATES TO MAINTAIN LIQUID LEVEL IN VARIOUS TANKS IN THE DUAL ALKALI PROCESS

Tank	Controlled Stream
Absorber	Absorber bleed
Primary reactor	Primary reactor overflow
Secondary reactor	Reactor bleed and/or secondary reactor overflow
Thickener	Thickener overflow
Thickener hold tank	Make-up water
Filter tub	Slurry overflow
Soda ash tank	Soda ash solution flow
Reactant feed tank	Lime slurry feed to reactant feed tank

the feed forward flow rate to the absorbers. The SO_2 concentration in the effluent gas from the absorber will be controlled to less than 200 ppm. The absorber liquor pH will be in the range of 5-6, corresponding to this SO_2 concentration in the absorber effluent flue gas. The exact correlation of outlet SO_2 with bleed pH will be developed during system startup.

Liquor from the absorber tank is recycled to the sprays and the trays. The feed forward from the thickener hold tank, controlled by the pH of the absorber bleed, is mixed with the absorber recycle liquor to the trays prior to entering the absorber. The recycle is used for pH control and to maintain adequate liquor flow across the trays. The pH of the mixed liquor stream is measured and is connected to an alarm system located on the control panel, so that if the pH drifts outside a preset range, the flow rates can be adjusted accordingly.

The absorber bleed is fed to the primary reactor along with lime slurry. The absorber bleed rate is controlled by liquid level in the absorber recycle tank. The lime slurry feed rate to each primary reactor is controlled by the pH of the secondary reactor liquor. The pH is a measure of the extent of reaction in the reactor system.

Carbide lime slurry is supplied to the primary reactors from the lime hold tank. The lime slurry is recycled around the slurry tank to maintain sufficient velocity in the pipes to avoid solids deposition. A bleed from the recycle is fed to the primary reactors and the excess lime slurry is returned to the lime hold tank. A control valve on the return lime slurry line is interconnected with the control valves on the lime slurry feed to the individual reactors. The recycle valve is used to maintain enough pressure in the feed lines to each primary reactor to ensure sufficient slurry feed and adequate feed control. If the lime slurry feed to a reactor is less than desired and the control valve on the lime slurry feed to this reactor is almost completely open, then the control valve on the lime slurry return line will be closed slightly so that the flow to the reactor is increased.

Slurry from the primary reactors overflows by gravity to the secondary reactors where the regeneration reaction is completed. The liquor level in the primary reactor is maintained by the height of the overflow weir. The liquor level in the secondary reactor is maintained by controlling the bleed rate from the reactor. The liquor level controls a valve on the discharge of the secondary reactor pumps. The slurry from the reactor is fed to the thickener center well.

The solids in the thickener are allowed to settle and the clear liquor overflows to the thickener hold tank. The underflow slurry from the thickener is pumped to the filters. In order to maintain sufficient velocity of the underflow slurry in the pipes to prevent deposition of solids, a constant recycle of slurry around the thickener is maintained from which a bleed is taken off to feed the filters. The recycled slurry is returned to the solids zone in the thickener.

The suspended solids concentration in the underflow slurry is a principal factor affecting the cake thickness and the washability of the cake. The slurry density is related to suspended solids concentration in the slurry and, therefore, the slurry density is controlled by the use of dilution liquor from the thickener hold tank and by the number of filters in service. The underflow slurry density is measured by a density meter. If the slurry is too thick, thickener hold tank liquor is added to dilute the slurry and/or an additional filter put on line. If the slurry is thin, dilution liquor is reduced or a filter shut down.

The slurry level in the filter tub is maintained by the position of the overflow weir. Excess slurry is fed to the filter to ensure the adequate level in the tub. Slurry overflowing the tub is returned to the thickener center well by gravity.

The filter cake is washed with makeup water to remove sodium salts in the occluded liquor in the cake. Combined filtrate (recovered liquor and wash water) is collected in the filtrate receiver and is returned to the thickener center well.

Any sodium value lost in the cake is replaced by adding an equivalent amount of the soda ash to the system. Process liquor from the thickener hold tank is pumped to the soda ash solution tank at a constant rate of 50 gpm. The soda ash is weighed on a weigh feeder and is fed to the soda ash solution tank. The soda ash solution is pumped to the thickener at a rate to maintain a constant level in the soda ash solution tank.

The soda ash feed rate is controlled by the total lime slurry feed rate to the two primary reactors. The lime slurry feed rate to each individual reactor is measured and summed, and the soda ash feed rate is set in direct proportion to the total lime slurry feed rate. The proportionality constant may be changed, if desired. This type of control is based upon the fact that the expected sodium loss varies proportional to the amount of lime fed to the system, and the soluble sodium lost in the cake is a function of the weight of the cake produced (for a constant wash ratio).

VII. OFFSITES AND AUXILIARIES

The offsites generally required for the dual alkali system include: services for electrical supply, water supply and instrument air; oil, steam or hot water if the wet flue gas is to be reheated; raw materials receiving and storage facilities; a wet chemicals analytical laboratory; and appropriate shop facilities for repair and maintenance of machinery and instruments. Except for electrical service, all of these offsites, including lime receiving and storage facilities, now exist at Cane Run Station and are available. An electrical substation including appropriate step-down transformers will be installed for the dual alkali system.

Since the system will use the same carbide lime as used in the existing direct lime scrubbing systems at Cane Run, additional lime receiving, storage and handling systems will not be required.

The offsites and services required for the Cane Run system are reviewed in the following sections. Significant differences between the requirements for the system at the Cane Run Station and general applications of dual alkali technology are indicated and alternative facilities or services discussed.

A. ELECTRICAL POWER

Electrical power available at Cane Run Station for construction and operation of the system include the following:

4,160 V ac, three-phase, 60 hertz
480 V ac, three-phase, 60 hertz
120 V ac, single-phase, 60 hertz

A feed line will be taken from the existing 14 KV substation and the voltage will be reduced in a step-down transformer to 4,160 volts. Step-down transformers for further voltage reduction to 480 volts and from 480 volts to 120 volts will also be installed by LG&E. The cost of this offsite (a feed line and the three step-down transformers) is included in the capital investment for the system.

The power requirement for the system is estimated at about 1.0% of the total power generated by the boiler at peak load. However, the design gas flow rate for the dual alkali system at LG&E is higher than the maximum flow rate and therefore the estimated electrical energy requirement should be conservative.

B. WATER SUPPLY

The maximum water requirement for the system is about 470 gpm, not including the water associated with the slurried lime feed. Of this, approximately 250 gpm is required for the process and about 220 gpm for noncontract cooling water.

River water will be used for all water requirements. The river water is available at the following conditions:

Water Supply Pressure	50-100 psig
Water Temperature	35-95°F
Total Dissolved Solids	300-500 ppm
Suspended Solids	50-500 ppm
pH	6-8
Na_2SO_4	20-200 ppm
Hardness	80-250 ppm
$CaCO_3$	50-250 ppm
Fe	0.1-30 ppm
Mn	0.15-2.5 ppm
NaCl	10-100 ppm

The water supply to the system will be filtered to prevent solids from entering the pump seals and filter spray nozzles. An inline filter will be used for this purpose, although the mesh size for the filter screen has not been selected.

Liquid waste streams from a plant could also conceivably be used as process makeup water for the dual alkali system. This offers the possibility of reducing the fresh water requirement as well as reducing effluent waste streams from the utility plant. Effects of chemicals present in the cooling water blowdown on dual alkali process chemistry, materials of construction, solids properties and raw material consumption have not been evaluated.

C. INSTRUMENT AIR

Air is available at Cane Run Station at 60-125 psig. Air is used only for instruments and air-operated controls. The total amount of air used for the process is small and existing excess compressor capacity at the plant should be adequate to supply the air. No new compressor capacity is anticipated.

D. OIL

No. 2 fuel oil will be used to reheat the wet exhaust gases from the absorbers. LG&E is planning to convert a 15 Mw gas turbine to an oil turbine and an oil tank with a 100,000 gallon capacity has been installed for this purpose. A separate pump will be used to supply oil from this tank to the dual alkali system. This same tank will also serve as the oil storage tank for other SO_2 removal systems at the Cane Run Station. The oil requirement for the dual alkali system to provide 50°F reheat at design load is estimated to be 343 gal./hr.

In general, oil, steam or hot water may be used for reheat; or in some cases, reheat may be supplied by bypassing part of the flue gas around the scrubber. The latter possibility exists only with systems capable of high SO_2 removal efficiencies (98 to 99% SO_2 removal) and thereby allowing reasonable overall

efficiencies even though part of the flue gas is not treated. If steam or hot water is used for reheat, no additional offsite equipment is required, since steam and hot water lines may be directly taken from the utility plant. Use of oil requires a storage tank and an oil pump; although in some cases, as in the system for Cane Run Station where oil is used as a fuel for electricity generation, storage of oil for reheat can be combined with the existing storage tanks at the plant.

E. CARBIDE LIME FACILITY

Carbide lime slurry will be used to regenerate the spent sodium solution. Since receiving, handling and storage facilities for the carbide lime slurry already exist at the plant, only a day tank to store lime slurry for the process will be installed as a part of the dual alkali system.

Carbide lime is available as slurry containing 30% dry solids. The slurry is shipped in the LG&E barges to the Cane Run Station. The slurry is pumped from the barge to an agitated storage tank from which it is pumped to a dual alkali day tank as required.

In general, calcined lime, hydrated lime, or carbide lime may be used to regenerate the spent sodium solution. While carbide lime is cheaper than commercial lime, it is not available at most locations. Normally, calcined lime would be used. It would be slaked and fed to the system as a slurry, and therefore would not be considered an offsite. It is included as an offsite here because the carbide lime facility already exists. Although capital investment for this facility is not included in the total estimated capital requirement for the dual alkali system at Cane Run, capital charges for the facility are included in the estimated cost of the carbide lime delivered to the system.

F. LABORATORY AND SHOP CAPABILITIES

The Cane Run Station has the necessary laboratory and shop capabilities for the maintenance and operation of the dual alkali system and no additional facilities are contemplated. The equipment required for wet chemical analyses is small and can usually be incorporated in the existing plant control lab.

1. Laboratory Capability

The laboratory equipment and materials for the chemical and physical testing required during the operation and testing of the dual alkali system are as follows:

- Analytical balances.
- Atomic absorption spectrophotometer.
- pH meter with electrodes for standard pH measurements and lead electrodes for sulfate titration.

- Forced draft-type oven with temperature control (± 0.5°C).
- Automatic burets and magnetic stirrers.
- Distilled water and various reagents for wet analyses.
- Assorted glassware for sample preparations.

Much of this equipment may already be part of an existing control laboratory at a power plant for use in monitoring and analysis of coal, cooling water, boiler feed water, and waste streams; or can be easily included as a part of the control laboratory equipment. In some cases, special analyses for metal ions (calcium, sodium, and magnesium) requiring the use of an atomic absorption spectrophotometer can be performed by outside testing laboratories.

2. Shop Capability

LG&E carries out their own plant construction. The Cane Run Station has adequate shop facilities to operate and maintain the boilers and the existing direct lime scrubbing systems. The shops are equipped with tools and equipment worth over $3.0 million, including a 175-ton capacity crane. No additional shop capacity or capabilities are anticipated.

In general, the required shop capacity for dual alkali systems include the following:

- Crane capacity to lift motor, pump, valves, etc., which need occasional maintenance.
- Machine shop to machine relatively simple surfaces, thread pipes, etc.
- Welding equipment, both in shop and in field.
- Instrument shop to check out instruments.
- Electrical shop.

VIII. CAPITAL AND OPERATING COSTS -- DUAL ALKALI SYSTEMS

The capital and operating costs for the dual alkali system are developed in this chapter. The estimated capital and operating costs are presented both for the dual alkali system installed on Cane Run Unit No. 6 and for generalized dual alkali systems for new 500 and 1,000 megawatt systems.

A. 300 Mw RETROFIT SYSTEM FOR LG&E CANE RUN NO. 6

1. Design Basis

The dual alkali system for Louisville Gas and Electric Company's Cane Run Station is a retrofit system installed on Unit No. 6, an existing boiler equipped with an electrostatic precipitator for particulate removal. The design basis for the system is given in Table VIII-1. The system is designed to handle the flue gas produced with the boiler operating at a gross peak load of 300 megawatts, and the design gas flow rate of 1,065,000 acfm at the scrubber inlet is about 10% higher than that estimated by McGlamary, et al. (1975) for an existing boiler having 300 Mw capacity. For coal containing 5.0% sulfur and 0.04% chloride (dry basis), the design SO_2 removal efficiency is 94.2%.

The SO_2 removal system will have two booster fans, two absorbers, two reactor modules, one thickener, and three filters. The system will be capable of operating in the range of 60 to 300 Mw boiler load. The waste exhaust gas from the absorber will be reheated by a direct oil fired reheater. The reheat is equivalent to 50°F. The system will have a bypass and appropriate spare capacity, as shown in Table VIII-1. The sludge will be disposed in an existing onsite pond after treatment. The cost of the pond is not included in the capital cost estimate.

Carbide lime will be used as a raw material to regenerate the sodium solution in the dual alkali process. Carbide lime is available from a local supplier. The use of carbide lime is expected to result in savings in the operating cost compared to the system based on commercial lime. The carbide lime will be supplied in a slurry form on a barge. An existing barge handling facility including a slurry storage tank will be used for the dual alkali system. At present, the carbide lime facility supplies lime for the two existing SO_2 removal systems installed at the Cane Run Station.

2. Capital Investment

The capital investment for the dual alkali demonstration system is shown in Table VIII-2. The capital costs are broken down into total materials, erection, engineering, spare parts and working capital. The overall project schedule is two years, with design starting in the latter part of 1976 and construction scheduled for completion in the latter part of 1978. The costs shown in Table VIII-2 are as incurred and therefore represent costs roughly equivalent to September 1977 dollars. The estimated capital investment for the system is $17,379,000, which corresponds to about $58.0/kilowatt (based on gross peak capacity).

TABLE VIII-1

DESIGN BASIS FOR THE DUAL ALKALI SYSTEM AT THE CANE RUN 6
LOUISVILLE GAS AND ELECTRIC COMPANY

Boiler Gross Peak Capacity, Mw	300
Boiler Gross Net Capacity, Mw	277
Annual Load Factor, %	60 (of gross peak capacity)
Btu/kwhr (basis-input heat/gross peak electricity generated)	9,960
S in Coal, %	5.0
Cl in Coal, %	0.04
Ash, %	17
Moisture, %	9
Btu/lb of Coal	11,000
Gas Flow Rate @ Scrubber Inlet, acfm	1,065,000
Temperature, °F	300
SO_2, lbs/min.	390
Fly Ash, lbs/min.	4.0
SO_2 @ Scrubber Inlet, lb/10^6 Btu	8.55
Land Cost	Not included
SO_2 Removal System:	Retrofit
SO_2 @ Scrubber Outlet, lb/10^6 Btu	0.50
SO_2 Removal in Scrubber, %	94.2
Particulate Removal	None
Bypass	Yes
Turndown	to 20% of Capacity
Number of Absorber/Reactor Modules	2
Number of Filter Units	3
Thickeners	1
Carbide Lime Feed System	Existing (new day tank)
Soda Ash Silo Capacity	14 days
Reheat	Direct Reheat with Oil-Fired Reheater
Sludge Disposal	Onsite Pond with Treatment
Spare Capacity,[a] %	
Filters	50
Pumps	100
Necessary Instruments to Run Process	100

[a]The system can be operated with one reactor system at full load and one absorber system at less than 60% load.

TABLE VIII-2

ESTIMATED CAPITAL COSTS FOR DUAL ALKALI SYSTEM AT THE CANE RUN 6
LOUISVILLE GAS AND ELECTRIC COMPANY

	Dollars
Total Materials Cost:	
Major Equipment Cost (Table VIII-3)	7,037,000
Other Materials Cost[a] (Table VII-4)	2,525,000
Sludge Disposal Equipment	900,000
Additive Slurry System[b]	700,000
Total Materials Cost	11,162,000
Erection:[d]	
Direct Labor (252,800 hrs. @ $12/hr)[c]	3,034,000
Field Supervision	273,000
Total Erection Cost	3,307,000
Engineering Costs:	
System Supplier Engineering	1,323,000
Owner's Engineering Expenses	303,000
Owner's Consulting Engineer	852,000
Total Engineering Cost	2,478,000
Spare Parts, 2% of the Total Materials Cost	232,000
Working Capital	200,000
Total Capital Investment[e]	17,379,000
$/kw (Based on 300 Mw gross peak load)	57.9
(Based on 277 Mw gross net load)	62.7

Basis: 300 Mw gross peak load existing coal fired boiler (277 Mw net peak load)
S in coal, 5.0%
SO_2 removal efficiency in scrubber, 94.2%
Stack gas reheat, 50% - direct oil fired reheater
Disposal of sludge after treatment to onsite pond
Project beginning mid 1976, ending late 1978, Avg. 1977 dollars
Necessary parts in storage and reasonable spare capacity

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[a] Sludge disposal equipment cost ($900,000) is shown separately. The $900,000 is the 3/7 portion of the total sludge disposal cost for Cane Run 4, 5 and 6.

[b] Additive supply system cost ($700,000) is shown separately. The $700,000 is the 3/7 portion of the total additive supply system for SO_x removal systems for Cane Run 4, 5 and 6.

[c] Includes plant overhead.

[d] Erection equipment cost is included in plant overhead.

[e] The capital investment for the above system in 1976 dollars is equivalent to $15.95 million or 53.15 $/kw (based on gross peak capacity).

The total materials costs include major equipment (including sludge disposal equipment), other materials (for offsites, foundations, etc.), and cost of the additive slurry system. The materials costs represent the delivered cost to the plant site including freight, insurance, taxes, and profit.

The breakdown of the cost of major equipment is shown in Table VIII-3. Major equipment includes absorbers, agitators, dampers, ductwork, expansion joints, fans, fluid drives, instrumentation, lining, motors, pumps, reheaters, structural steel, tanks, thickener shell and mechanism, vacuum filters, valves and piping, soda ash silo and auxiliaries. The cost of the major equipment is $6.2 million in January 1976 dollars. It is expected that most of the items will be delivered in mid-1977. The total, escalated cost at 13.5% per 1-1/2 years (to September 1977) is $7.03 million.

The breakdown of the other materials cost is given in Table VIII-4. These costs include materials for buildings, land improvement, insulation, and foundation for major equipment, and materials requirements for offsites. Also, costs for the chimney liner, breeching, and accessories, and waste disposal equipment are included as a part of additional materials. The total cost of the purchased parts is $3.4 million. This is in addition to the cost of major equipment. Again, the freight charges, taxes, insurance, startup and modifications have been included in the other materials cost and are not shown separately. The costs are escalated to the estimated dates of delivery of materials.

Chimney liner coating and breeching and accessories represent a direct retrofit cost. The total cost for these items as shown in Table VIII-4 is $400,000. The electrical offsite cost is $774,900 and includes station control, feedline for 14 KV, appropriate stepdown transformers to reduce voltage to 4160 V, 480 V and 120 V, and grounding connections.

The waste disposal cost of $900,000 represents 3/7 of the total estimated capital investment for the waste disposal system for all three scrubbing systems at the Cane Run Station. The 3/7 factor is equal to the capacity ratio for the Cane Run Unit No. 6 and the combined capacities of Cane Run Units Nos. 4, 5, and 6. Similarly, the additive slurry system cost of $700,000 represents 3/7 of the total estimated capital investment for the total additive slurry system for all three scrubbing systems at the Cane Run Station.

The cost of land for the sludge disposal pond or for the SO_2 removal system is not included in the capital cost, as the system represents retrofit situation and, at present, no expenses are incurred.

The capital requirement for the carbide lime handling facility and the storage facility is not included in Table VIII-2. However, the capital investment for the day tank for storage of carbide lime slurry used in the dual alkali system is included in the total cost shown in Table VIII-2. Similarly, the initial cost of additional capacities for compressor pump to supply instrument air, service water pump to supply makeup water, oil storage

TABLE VIII-3

COST OF MAJOR EQUIPMENT COMPONENTS FOR DUAL ALKALI SYSTEM AT THE LOUISVILLE GAS AND ELECTRIC COMPANY

Equipment	Cost[a]
Absorbers	778,165
Agitators	82,783
Dampers	310,986
Ductwork	437,471
Expansion Joints	139,497
Fans	367,770
Fluid Drives	124,457
Instrumentation	240,898
Lining	366,626
Motors	516,969
Pumps	204,058
Reheaters	144,739
Structural Steel	279,276
Tanks	526,198
Thickener Mechanism	171,093
Thickener Shell	122,553
Vacuum Filters	838,404
Valves and Piping	505,364
Weigh Feeder	43,082
	6,200,389
Escalation cost for 13 1/2% for 1 1/2 year period	837,052
Total	7,037,441

[a]The equipment cost includes freight, insurance, taxes, etc. and is delivered cost to Louisville Gas and Electric Company.

TABLE VIII-4

ADDITIONAL PURCHASED PARTS COST FOR DUAL ALKALI SYSTEM
AT THE LOUISVILLE GAS AND ELECTRIC COMPANY

Pump and Filter Buildings	141,125
Equipment for Land Improvements	24,500
Insulation and Foundation for Fan	40,900
Foundation, etc. for Major Components	703,400
Instrumentation and Test Equipment	20,000
Concrete Ducts	18,400
Chimney Liner Coating[a]	288,000
Breeching and Accessories[a]	112,000
Waste Disposal Equipment	900,000
Electrical Offsite	
Station Control[b]	206,900
14KV OCB and Equipment[b]	45,500
4160V Auxiliary[b]	370,000
460V Auxiliary[b]	145,100
Station Grounding[b]	7,400
Piping for Services	300,900
Indirect Materials	93,000
Taxes	8,300
TOTAL	3,425,425

Note: Purchased parts cost does not include cost of major equipment components shown in Table VIII-3. The purchased parts cost includes freight, etc. and is delivered cost to Louisville Gas and Electric Company.

[a] Costs directly related to retrofit - $400,000.

[b] Electrical utilities - $774,900.

tank, and switch gear building are not included in the estimated capital investment.

The erection labor shown in Table VIII-2 includes electricians, boilermakers, pipefitters, carpenters, welders, etc., and covers the erection of major equipment, temporary facilities, and plant startup and modifications. Because of the low labor rates at this plant, the average escalated labor cost including the overhead is $12 per hour. Including field supervision, the total erection cost is expected to be $3.3 million.

The engineering costs include: engineering by system supplier, covering selection of major equipment and design of the dual alkali process; consulting engineering to provide the remainder of the engineering not covered by system supplier; and local engineering and expenses representing owner's cost. The total engineering cost is approximately $2.5 million and is equal to 17% of the total material and erection costs.

The cost of spare parts is assumed to be 2% of the total materials cost. Startup and modification costs are included in equipment and labor costs and are not broken out as a separate cost item. Working capital is assumed to be $200,000.

3. Annual Operating Costs

The system is expected to start up in late 1978, and therefore 1979 will be the first full year of operation. The estimated annual operating costs for the dual alkali system in 1979 dollars are shown in Table VIII-5.

The estimated annual operating costs in 1976 dollars have also been developed, since these represent more well-defined costs using recent unit cost data. Costs in 1976 dollars are shown in Table VIII-6.

Operating costs are based on an average sulfur content in the coal of 3.8%. The average sulfur dioxide removal rate is assumed to be 297 lbs/minute (94.2% SO_2 removal efficiency) at 300 Mw boiler load. The operating costs are estimated both with and without operation of the reheat system. When included, reheat is assumed to be 50 F°. The sludge disposal costs are based on onsite solids disposal after mixing with ash and lime.

The total annual operating cost without reheat is estimated to be $3.6 million in 1976 and $4.3 million in 1979. The corresponding costs with reheat are about 15% higher. The 1976 operating costs range from 2.3 mills/kwhr (23.0¢ per 10^6 Btu) without reheat system to 2.7 mills/kwhr (26.7¢ per 10^6 Btu) with 50 F° reheat. The cost estimates for the year 1979 are about 20% higher.

The estimated cost of carbide lime in 1979 is $13.29/ton. This includes: (a) cost of lime at $2.00/ton; (b) handling cost to barges at $5.53/ton; (c) transportation in barge to Cane Run at $1.21/ton; (d) cost to unload barge and supply lime slurry to dual alkali system at $4.55/ton. These costs include capital charges and depreciation on capital investment.

TABLE VIII-5

ESTIMATED AVERAGE ANNUAL OPERATING COSTS
FOR DUAL ALKALI AT THE LOUISVILLE GAS AND ELECTRIC COMPANY
(1979 $)

	Quantity	Unit Cost, $	Total Costs, $
Direct Costs:			
Carbide Lime	58,728 tons	13.29/ton	780,500
Soda Ash	1,912 tons	78.65/ton	150,400
Fuel Oil[a]	1,802,808 gals.	0.43/gal.	775,200
Electricity	16,188,480 kwhrs	0.01/kwhr	161,900
Water	126,100,000 gals.	0.05/1,000 gals.	6,300
Sludge Removal	185,280 tons	2.01/wet ton	372,400
Maintenance Materials	2.5% of total materials cost		279,000
Labor			
Operation	26,280 hrs	8.18/hr	215,000
Maintenance	26,280 hrs	8.28/hr	217,600
Analysis	2,080 hrs	10.00/hr	20,800
Supervision	2,080 hrs	Various	40,000
Total Direct Costs			3,019,100
		No Reheat[a]	2,243,900
Indirect Costs:			
Overhead	59.4% of 493,368 (Total Labor)		293,100
Interest	6.125% of 17,379,000		1,064,500
Depreciation	4.17% of 17,379,000		724,700
Total Indirect Costs			2,082,300

	With Reheat	[a]No Reheat
Total Annual Operating Cost, $	5,101,400	4,326,200
Mills/kwh	3.24	2.74
¢/10^6 Btu	32.5	27.6
$/Ton of S Removed	217.9	184.7

Basis: 300 Mw (gross peak capacity) existing coal fired plant, 60% load factor, 9,960 Btu/kwhr, 3.8% S in coal 94.2% SO_2 removal, average SO_2 removed - 297 lbs/min. at 300 Mw (gross peak) load, on-site solids disposal by trucking after treatment, stack gas reheat 50°F.

TABLE VIII-6

ESTIMATED AVERAGE ANNUAL OPERATING COSTS FOR DUAL ALKALI SYSTEM
AT THE LOUISVILLE GAS AND ELECTRIC COMPANY
(1976 Dollars)

	Quantity	Unit Cost, $	Total Costs, $
Direct Costs:			
Carbide Lime (30% slurry)	58,728 tons	9.91/ton	582,000
Soda Ash	1,912 tons	65.00/ton	124,300
Fuel Oil[a]	1,802,808 gals.	0.32/gal.	576,900
Electricity	16,188,480 kwhrs	0.01/kwhr	161,900
Water	126,100 000 gals.	0.05/1,000 gals.	6,300
Sludge Removal	185,280 tons	1.51/wet ton	279,800
Maintenance Materials	2.5% of total materials cost[a]		256,000
Labor			
Operation	26,280 hrs	6.00/hr	157,700
Maintenance	26,280 hrs	6.00/hr	157,700
Analysis	2,080 hrs	8.00/hr	16,600
Supervision	2,080 hrs	12.00/hr	25,000
Total Direct Costs			2,344,200
		[a]No Reheat	1,767,300
Indirect Costs:			
Overhead	55.8% of 356,960 (Total Labor)		199,200
Interest	6.125% of 15,950,000		976,900
Depreciation	4.17% of 15,950,000		665,100
Total Indirect Costs			1,841,200

	With Reheat	[a]No Reheat
Total Annual Operating Cost, $	4,185,400	3,608,500
Mills/kwhr	2.65	2.29
¢/10^6 Btu	26.7	23.0
$/Ton of S Removed	178.8	154.1

Basis: 300 Mw (gross peak load) existing coal fired plant
60% load factor
9,960 Btu/kwhr
3.8% S in coal
94.2% SO_2 removal
Average SO_2 removal - 297 lbs/min. at 300 Mw.
On-site solids disposal by trucking after treatment
Stack gas reheat - 50°F

[a]Total materials cost of $11.16 million in 1977 dollars is equivalent to $10.24 million in 1976 dollars.

The delivered cost of the dense soda ash to the silo is estimated to be $65/ton in 1976 and $78.65/ton in 1979. The fuel oil cost is estimated at $0.32/gal. in 1976 and $0.43/gal. in 1979. The cost of electricity for internal use at LG&E is $0.01/kwhr. The water cost is estimated to be $0.05/1,000 gals. for pumping river water.

The cost for disposal of filter cake is estimated to be $1.51/ton (wet basis) in 1976 and $2.01/ton (wet basis) in 1979, assuming 55% solids in the cake. The disposal cost is estimated based on trucking the filter cake to the onsite pond. The estimated manpower requirement for the disposal operation is 5 men/shift.

The system operating and maintenance requirements are estimated to be on the average 3 men/shift for each; the materials cost for maintenance is estimated to be 2.5% of the total materials cost. The analysis and supervision is estimated to be 1 shift/day of effort for each.

The indirect costs include overhead, interest, and depreciation. Current overhead rate of 55.8% for LG&E is used for the operating cost estimate in 1976 and the expected overhead rate of 59.4% is used for the operating cost estimate in the year 1979. The overhead rate is applicable to all labor charges only. Based on the interest rate of 6.125% on the nontaxable bonds floated by the LG&E, interest cost is estimated at about $1.06 million in 1979 and $0.976 million in 1976. Depreciation is estimated based on the straight line method using an estimated life of 24 years for the flue gas desulfurization system. For estimating capital charges in the year 1979, 1977 dollars are used because LG&E does not capitalize interest cost.

B. GENERALIZED CAPITAL AND OPERATING COSTS

1. Description of Systems

In this section, generalized costs are developed for dual alkali systems operating in the concentrated mode using lime for regeneration. The costs are estimated for applications to new 500 and 1,000 megawatt boilers. Sufficient detail is provided to allow estimates of other size systems and for comparisons with costs of other SO_2 removal systems.

The process consists of: an SO_2 scrubbing system including fans and bypass ducting with appropriate controls for reheating the wet flue gas with a bypassed gas stream; a reactor system; a thickener and filter system for solids separation; filter cake handling equipment and disposal pond; lime and soda ash receiving, handling and feed preparation equipment, including lime slaking; and required offsites. Particle removal may be incorporated in the dual alkali system; however, it is not included here. Flue gas from the electrostatic precipitator passes through booster fans and then either to the scrubbers or via a bypass to the stack. The booster fans for the dual alkali system also serve as induced draft fans for the boiler. In the SO_2 scrubber system, the flue gas passes through the absorption section and a demister. The wet flue gas is then mixed with bypass gas before being discharged to the stack.

Liquid is recycled in the absorbers to maintain the required L/G ratio and a bleed stream consisting of spent absorber solution is sent to the reactor system. Hydrated lime slurry is used to regenerate the spent sodium solution. The reactor bleed is a slurry containing calcium solids and a regenerated solution of sodium salts. The slurry is thickened in the thickener and the thickener underflow slurry is filtered to separate the calcium sulfur salts. The cake is washed to recover the sodium salts in the entrained solution and the combined filtrate (recovered liquor and wash water) is returned to the thickener. The cake is trucked to an onsite disposal pond one mile from the plant. Clear, regenerated liquor overflows the thickener to the thickener hold tank from which it is fed to the absorber where it is mixed with the absorber recycle stream.

Calcined lime is received and stored in silos having 15 days' storage capacity (design basis). The lime is slaked and the slurry stored in a day tank from which slurry is supplied to the reactor system. Dry soda ash added to the system to replace sodium value lost in the washed cake, is stored in a silo having a 15-day capacity. The soda ash is added directly to the process liquor in a soda ash solution tank from which it is pumped to the thickener.

In order to provide a common basis for comparison of generalized system costs, some of the assumptions used by McGlamery (1975) in prior studies of the economics of SO_2 scrubbing have been incorporated in the capital and operating cost estimates. The basis for the costs is coal containing 3.5% sulfur and 0.1% chloride and having a heating value of 12,000 Btu/lb. Three cases have been considered: (1) a new 500 megawatt boiler with SO_2 removal necessary to meet current Federal New Source Performance Standards (NSPS); (2) a new 500 megawatt boiler with 90% SO_2 removal; and (3) a new 1,000 megawatt boiler with SO_2 removal to meet Federal NSPS.

The design bases for the three cases are shown in Table VIII-7. The systems are assumed to be modular in nature. There are three scrubber/reactor trains in the 500 Mw system and six scrubber/reactor trains in the 1,000 Mw system.

As previously indicated, the systems are equipped with flue gas bypass, and reheat for the wet gas is provided by a bypass gas stream. The SO_2 removal efficiencies indicated in Table VIII-7 are the overall SO_2 removal efficiencies. Removal efficiencies for the flue gas passing through the absorber may be 98% or higher. In the cases where the overall SO_2 removal is 78.1% to comply with NSPS, the amount of flue gas bypassed is 20% and the amount of reheat provided is 35°F. In the cases where overall SO_2 removal is 90%, the amount of flue gas bypassed is 8% and the reheat provided is 14°F.

While reheat is provided with bypassed gas, absorbers are designed to handle 100% of the boiler flue gas. Therefore, the 1,000 system designed to meet NSPS should be able to operate at full load with one absorber shut down, and the 500 Mw systems should be able to operate at 75-80% load with one absorber out of operation.

TABLE VIII-7

DESIGN BASIS FOR THE GENERALIZED DUAL ALKALI SYSTEM

Boiler Capacity, Mw	500	500	1,000
Boiler Location	Midwest	Midwest	Midwest
Boiler Status	New	New	New
Coal Heating Value, Btu/lb	12,000	12,000	12,000
Btu/kwhr	9,000	9,000	9,000
Boiler Operating Time, hrs/yr	7,000	7,000	7,000
Coal Burned, tons/yr	1,312,500	1,312,500	2,625,000
S in Coal, %	3.5	3.5	3.5
Cl in Coal, %	0.10	0.10	0.10
SO_2 in Flue Gas (94% of S in coal), lb/hr	24,675	24,675	49,350
SO_2 @ Scrubber Inlet, lb/10^6 Btu	5.48	5.48	5.48
SO_2 @ Stack, lb/10^6 Btu	1.2 (NSPS)	0.548	1.2 (NSPS)
SO_2 Removal	78.1%	90%	78.1%
lbs/hr	19,270	22,200	38,540
tons/yr	67,450	77,725	134,900
Gas Flow Rate @ Scrubber Inlet, acfm	1,540,000	1,540,000	3,080,000
Temperature, °F Scrubber Inlet, acfm	310	310	310
Flue Gas Oxygen, %	6	6	6
Oxidation in Dual Alkali System, %	10	10	10
Boiler I.D. Fan and Scrubber F.D. Fan	Common	Common	Common
Bypass	Yes	Yes	Yes
Flue Gas Bypassed for Reheat, %	20	8	20
Reheat Provided by Bypass Flue Gas, °F	35	14	35
Base Year for Cost Estimation	1977	1977	1977
Scrubber/Reactor Trains	3	3	6
Spare Capacity			
Absorbers, %	25	9	25
Filters, %	50	50	25
Pumps, %	100	100	100
Instruments, Necessary for Operation, %	100	100	100
Turndown to % of capacity	20	20	20
Sludge Disposal	Onsite	Onsite	Onsite
Available CaO in Raw Lime, %	92.5	92.5	92.5
Lime Utilization, %	98	98	98
Lime Silo Capacity, days	15	15	15
Soda Ash Silo Capacity, days	15	15	15

Each system can be operated at 100% load with one reactor system out of operation. Similarly, each filter can be operated independently of the others, and the number of filters in operation will therefore depend on the system load. Due to the modular nature of the filters, 50% spare filter capacity is provided for the 500 megawatt systems, and 25% spare capacity is provided for the 1,000 megawatt system. In all cases, one spare filter is provided.

The location selected to estimate labor costs is the Midwest. The project is assumed to start in mid-1976 and end in late 1978. Therefore, it is on the same basis as the dual alkali project at LG&E, and the capital costs can be assumed to be in September 1977 dollars.

2. Capital Investment

The capital costs for the three cases are shown in Table VIII-8. The costs are developed in parallel with the LG&E costs. The major equipment cost was obtained by using the costs shown in Table VIII-3 for the LG&E system and applying cost factors shown in Table VIII-9. Due to the modular nature of the system, the sizes of the various pieces of equipment do not vary significantly, and, therefore, application of cost factors results in minimal error. The cost of lining was assumed to vary directly with the cost of tanks, thickener, etc., which are lined. The cost of instruments was assumed to be proportional to the number of modules. The cost of the remaining equipment was assumed to vary accordingly to the average size cost factor.

Common fans are used for the dual alkali system and the boiler. One fan is provided for each absorber. The fan cost is the incremental cost over that for the induced draft fan required for the operation of the boilers. The incremental cost is equal to the difference in the costs of the fans having pressure drops of 27" WG and 15" WG.

The cost of lime handling and preparation facility includes: storage silos (eight silos having a total of 15-day storage capacity); two slakers; and the necessary equipment and instrumentation for conveying dry lime, filtering exhaust air from bins, vibrating silo bottoms, weighing and feeding, and pumping lime slurry. These costs were obtained from the TVA report by McGlamery, et al., prepared in 1975, and costs are updated to 1977 dollars.

Other equipment costs include materials for building, land improvements, insulation, and foundations, and materials required for offsites and auxiliaries. The costs of chimney liners and breeching, which are included in the LG&E system because of the retrofit situation, are not included here.

The costs of the onsite sludge disposal ponds shown in Table VIII-8 include the cost of land, construction of ponds and roads and accessways. The land required for the sludge pond and the flue gas desulfurization system is assumed to be 175 and 200 acres for the 500 megawatt systems having 78.1% and 90% SO_2 removal efficiency, respectively; and 350 acres for the 1,000 megawatt system. A cost of land equal to $3,000/acre is used. The cost of

TABLE VIII-8

ESTIMATED CAPITAL INVESTMENT FOR GENERALIZED DUAL ALKALI SYSTEM

Boiler Capacity SO_2 Removal Efficiency	500 Mw 78.1%	500 Mw 90%	1000 Mw 78.1%
Direct Costs	Dollars	Dollars	Dollars
Major Equipment Cost	9,970,000	10,488,000	17,397,000
Other Materials Cost	3,002,000	3,159,000	5,240,000
Erection Labor	6,486,000	6,823,000	11,319,000
Total Direct Costs	19,458,000	20,470,000	33,956,000
Indirect Costs			
Engineering Design (9% & 8%)	1,751,000	1,842,000	2,716,000
Construction Field Expense (11% & 10%)	2,140,000	2,252,000	3,396,000
Contractor's Fee (5%)	973,000	1,024,000	1,698,000
Contingency (10%)	1,946,000	2,047,000	3,396,000
Total Indirect Costs	6,810,000	7,165,000	11,206,000
Total Direct and Indirect Costs	26,268,000	27,635,000	45,162,000
Allowance for Spare Parts (2% of Equipment Cost)	259,000	273,000	453,000
Land	525,000	600,000	1,050,000
Working Capital	300,000	300,000	500,000
Total Capital Investment[a]	27,352,000	28,808,000	47,165,000
$/kw[a]	54.7	57.6	47.2

Basis: New coal fired boiler
3.5% S and 0.1% Cl in coal
Stack gas reheat by bypassing the flue gas
Sludge disposal by trucking to onsite pond (included in operating costs)
Midwest plant location, project beginning mid-1976 and ending late 1978, Sept. 1977 dollars
Necessary spare parts in storage, working capital included and reasonable spare capacity
Construction labor shortage with accompanying overtime pay incentive not considered

[a]The total capital investment for the above three systems in 1976 dollars is equivalent to $27,491,000, $29,207,000 and $47,977,000. These costs reported as $/kw in 1976 dollars are 54.98, 58.41 and 47.98, respectively.

TABLE VIII-9

SIZE COST FACTORS USED TO ESTIMATE CAPITAL COSTS

	Size Cost Scale Factor[a]
Absorbers	0.60
Agitators	0.50
Conveyor Belts	0.65
Conveyor Elevator	0.65
Dampers	0.65
Expansion Joints	0.65
Fans	0.68
Feeder Bin	0.58
Fluid Drives	0.68
Reheaters	0.80
Slaker	0.57
Storage Bins	0.68
Tanks	0.68
Thickener Mechanism	0.50
Thickener Shell	0.68
Vacuum Filters	0.65
Weigh Feeder	0.58

[a]Cost factors are applied as follows:

$$\frac{\text{Cost for Size A equipment}}{\text{Cost for Size B equipment}} = \left(\frac{\text{Size A}}{\text{Size B}}\right)^{\text{Cost factor}}$$

a general contractor to construct the sludge ponds, the access roads, and related dirt-moving tasks is estimated at $1.00/cubic yard in 1973 (Rossoff and Rossi, 1974) and escalated to $1.80/cubic yard for 1977 ($2.64/wet ton). The ponds are 40-feet deep and are assumed to be clay lined (McGlamery, 1975). Pond construction is assumed to proceed in conjunction with plant operation and is included as a part of operating costs.

The direct labor cost for erection of the dual alkali equipment is assumed to be proportional to the cost of equipment plus other materials. Since the erection labor at LG&E is low (only 32% of the equipment and materials costs), a higher value is assumed for generalized cost estimates (50% of the equipment and materials cost) consistent with the estimates given by McGlamery, et al., (1975). The erection labor includes labor requirements for temporary field construction and startup and modifications.

The total direct costs are $19.5-20.5 million for the 500 megawatt systems and $34.0 million for the 1,000 megawatt system. The engineering design costs are estimated at 9% of direct cost for the 500 megawatt systems and 8% of direct cost for the 1,000 megawatt system, and construction field expense is estimated at 11% for the 500 megawatt systems and 10% for a 1,000 megawatt system. The slightly lower relative engineering and field expenses for the 1,000 megawatt system reflect the modular nature of the systems and normal capacity factor cost savings.

Contractor's fees and contingency are taken at 5% and 10% of direct costs, respectively. Other costs are allowance for spare parts at 2% of equipment and materials cost, and working capital at $300,000 for the 500 Mw systems and at $500,000 for 1,000 Mw system.

The total capital costs in September 1977 dollars for the 500 megawatt systems are about $27.3 and $28.8 million, and about $47.2 million for the 1,000 megawatt system. The costs are equivalent to a range of $47-58/kw.

3. Operating Costs

The operating costs for the generalized systems are estimated in 1976 dollars and are shown in Tables VIII-10 to VIII-12. The costs are developed in a manner similar to those for the system at Cane Run Station. The differences between the LG&E system and generalized system are: (1) commercial lime is used instead of the carbide lime (the cost of lime delivered in the silo is assumed to be $40/ton); (2) the cost of electricity is estimated at $0.02/kWh instead of $0.01/kwhr used for the LG&E system; (3) the fuel cost is eliminated, since reheat is provided by flue gas bypassed; (4) the labor costs are assumed to be higher, because of the lower labor rates at the LG&E; and (5) capital charges are assumed to be 14.6% of the capital investment consistent with accounting practices in the industry.

TABLE VIII-10

ESTIMATED ANNUAL OPERATING COSTS FOR DUAL ALKALI SYSTEM (1976 Dollars)

500 Mw Boiler, SO_2 Removal to Meet NSPS (78.1% Removal Efficiency)

	Annual Quantity	Unit Cost, $	Total Cost, $
Direct Costs			
Raw Materials			
Lime	65,224 tons	40.00/ton	2,609,000
Soda Ash	3,039 tons	65.00/ton	198,000
Subtotal Raw Materials			2,807,000
Conversion Costs			
Operating Labor	26,280 hrs	8.00/hr	210,200
Analysis	2,080 hrs	10.00/hr	20,800
Supervision	2,080 hrs	12.00/hr	25,000
Utilities:			
Electricity	38,900,000 kwh	0.02/kwh	778,000
Water	420 x 10^6 gal	0.05/1,000 gal	21,000
Sludge Disposal[a]	227,700 wet tons	4.25/ton[a]	967,700
Maintenance	5% of direct investment[b]		892,600
Subtotal Conversion Costs			2,915,300
Subtotal Direct Costs			5,722,300
Indirect Costs			
Capital Charges	14.6% of fixed plant investment[c]		3,553,200
Overhead			
Plant	50% of conversion costs less utilities		1,058,200
Administrative	10% of operating labor		21,000
Subtotal Indirect Costs			4,632,400
Total Annual Operating Cost			10,354,700
Mills/kwh			2.96
¢/10^6 Btu			32.9
$/ton of Sulfur Removed			307

Basis: Operating time - 7,000 hrs/yr
9,000 Btu/kwh
3.5% S in coal
Average SO_2 removal - 321 lbs/min
Onsite solids disposal by trucking
Reheat 35°F

[a] Sludge disposal costs in 1976 dollars (per ton of wet sludge) = $2.75 for pond construction (including engineering, permits, contingency, etc. @ 14%) + $1.50 for sludge transport and placement.
[b] Direct investment of $19,458,000 in 1977 dollars is equivalent to $17,851,000 in 1976 dollars.
[c] Fixed plant investment of $26,527,000 in 1977 dollars is equivalent to $24,337,000 in 1976 dollars.

TABLE VIII-11

ESTIMATED ANNUAL OPERATING COSTS FOR DUAL ALKALI SYSTEM (1976 Dollars)

500 Mw Boiler, 90% SO_2 Removal

	Annual Quantity	Unit Cost, $	Total Cost, $
Direct Costs			
Raw Materials			
Lime	75,550 tons	40.00/ton	3,022,000
Soda Ash	3,368 tons	65.00/ton	218,900
Subtotal Raw Materials			3,240,900
Conversion Costs			
Operating Labor	26,280 hrs	8.00/hr	210,200
Analysis	2,080 hrs	10.00/hr	20,800
Supervision	2,080 hrs	12.00/hr	25,000
Utilities:			
Electricity	38.9×10^6 kwh	0.02/kwh	778,000
Water	420×10^6 gals	0.05/1,000 gal	21,000
Sludge Disposal[a]	263,750 wet tons	4.25/wet ton[a]	1,120,900
Maintenance	5% of direct investment[b]		939,000
Subtotal Conversion Costs			3,114,900
Subtotal Direct Costs			6,355,800
Indirect Costs			
Capital Charges	14.6% of fixed plant investment[c]		3,738,200
Overhead			
Plant	50% of conversion costs less utilities		1,158,000
Administrative	10% of operating labor		21,000
Subtotal Indirect Costs			4,917,200
Total Annual Operating Cost			11,273,000
Mills/kwh			3.22
¢/10^6 Btu			35.8
$/ton of Sulfur Removed			290

Basis: Operating time - 7,000 hrs/yr
9,000 Btu/kwh
3.5% S in coal
Average SO_2 removal - 370 lbs/min
Onsite solids disposal by trucking
Reheat 18°F

[a] Sludge disposal costs in 1976 dollars (per ton of wet sludge) = $2.75 for pond construction (including engineering, permits, contingency, etc. @ 14%) + $1.50 for sludge transport and placement.
[b] Direct investment of $20,470,000 in 1977 dollars is equvalent to $18,780,000 in 1976 dollars.
[c] Fixed plant investment of $27,908,000 in 1977 dollars is equivalent to $25,604,000 in 1976 dollars.

TABLE VIII-12

ESTIMATED ANNUAL OPERATING COSTS FOR DUAL ALKALI SYSTEM (1976 Dollars)

1,000 Mw Boiler, SO_2 Removal to Meet NSPS (78.1% Removal Efficiency)

	Annual Quantity	Unit Cost, $	Total Cost, $
Direct Costs			
Raw Materials			
Lime	130,448 tons	40.00/ton	5,218,000
Soda Ash	6,078 tons	65.00/ton	395,100
Subtotal Raw Materials			5,613,100
Conversion Costs			
Operating Labor	35,040 hrs	8.00/hr	280,300
Analysis	2,080 hrs	10.00/hr	20,800
Supervision	2,080 hrs	12.00/hr	25,000
Utilities:			
Electricity	77.8×10^6 kwh	0.02/kwh	1,556,000
Water	840×10^6 gals	0.05/1,000 gal	42,000
Sludge Disposal[a]	455,400 wet tons	4.25/wet ton[a]	1,935,500
Maintenance	5% of direct investment[b]		1,557,600
Subtotal Conversion Costs			5,417,200
Subtotal Direct Costs			11,030,300
Indirect Costs			
Capital Charges	14.6% of fixed plant investment[c]		6,110,000
Overhead			
Plant	50% of conversion costs less utilities		1,909,600
Administrative	10% of operating labor		28,000
Subtotal Indirect Costs			8,047,600
Total Annual Operating Costs			19,077,900
Mills/kwh			2.73
¢/10^6 Btu			30.3
$/ton of Sulfur Removed			283

Basis: Operating time - 7,000 hrs/yr
9,000 Btu/kwh
3.5% S in coal

Average SO_2 removal - 642 lbs/min
Onsite solids disposal by trucking
Reheat 35°F

[a] Sludge disposal costs in 1976 dollars (per ton of wet sludge) = $2.75 for pond construction (including engineering, permits, contingency, etc. @ 14%) + $1.50 for sludge transport and placement.

[b] Direct investment of $33,956,000 in 1977 dollars is equivalent to $31,152,000 in 1976 dollars.

[c] Fixed plant investment of $45,615,000 in 1977 dollars is equivalent to $41,849,000 in 1976 dollars.

The average labor is estimated at three operators/shift and four maintenance personnel/shift for the 500 megawatt systems and five operators/shift and six maintenance personnel/shift for the 1,000 megawatt system. It is expected that maintenance will be heavy on weekdays and light on evenings, nights and weekends. In addition, the cost of one analyst and one supervisor is included. The plant overhead rate is estimated to be 50% of the conversion costs less utilities. Capital charges, including interest, depreciation, taxes, insurance, and return on investment, are assumed to be 14.6% of the total capital investment.

The annual operating costs are $10.4 and $11.3 million for the 500 megawatt systems and $19.1 million for the 1,000 megawatt system. These costs are equivalent to a range of 30¢ to 36¢ per 10^6 Btu (2.7-3.2 mills/kwhr).

IX. REFERENCES

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4. Guthrie, K.M., "Capital Cost Estimating," Chemical Engineering, Volume 76, No. 6., March 24, 1969, pp. 114-142.

5. Rossoff, J. and Rossi, R.C., "Disposal of By-products from Non-regenerable Flue Gas Desulfurization Systems," Initial Report, EPA Contract 650/2-74-037A.

X. GLOSSARY

Active Sodium - Sodium associated with anions involved in SO_2 absorption reactions and includes sulfite, bisulfite, hydroxide and carbonate/bicarbonate. Total active sodium concentration is calculated as follows:

$$[Na^+]_{active} = 2 \times ([Na_2SO_3] + [Na_2CO_3]) = [NaHSO_3] + [NaOH] + [NaHCO_3]$$

Available Alkali - The percentage of the calcium hydroxide in the raw hydrated lime, or in the insoluble solids in the carbide lime slurry.

Calcium Utilization - The percentage of the calcium in the lime or limestone which is present in the solid product as a calcium-sulfur salt. Calcium utilization is defined as:

$$\text{Calcium Utilization} = \frac{\text{mols } (CaSO_3 + CaSO_4) \text{ generated}}{\text{mol Ca fed}} \times 100\%$$

Concentrated Dual Alkali Modes - Modes of operation of the dual alkali process in which regeneration reactions produce solids containing $CaSO_3 \cdot 1/2\ H_2O$ or a mixed crystal containing calcium sulfite and calcium sulfate hemihydrates, but not containing gypsum. Active sodium concentrations are usually higher than 0.15M Na^+ in concentrated mode solutions.

Dilute Dual Alkali Modes - Modes of operation of the dual alkali process in which regeneration reactions produce solids containing gypsum ($CaSO_4 \cdot 2\ H_2O$). Active sodium concentrations are usually lower than 0.15M Na^+ in dilute mode solutions.

Sulfate Formation - The oxidation of sulfur (IV)-sulfite and bisulfate. The level of sulfate formation relative to SO_2 absorption is given by:

$$\text{Sulfate Formation} = \frac{\text{mols S(IV) oxidized}}{\text{mol } SO_2 \text{ removed}} \times 100\%$$

Sulfate Precipitation - The formation of $CaSO_4 \cdot XH_2O$ insoluble solids. The level of sulfate precipitation in the overall scheme is given by the ratio of calcium sulfate to the total calcium-sulfur salts produced:

$$\text{Sulfate Precipitation} = \frac{\text{mols } CaSO_4 \text{ produced}}{\text{mols } (CaSO_3 + CaSO_4) \text{ produced}}$$

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